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FLOC STRENGTH AND FILTERABILITY OF PRETREATED WATER

Prepared For

FEDERAL WATER POLLUTION CONTROL ADMINISTRATION
DEPARTMENT OF INTERIOR

OCTOBER 1968

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LUDWIG ENGINEERING AND SCIENCE RESEARCH FOUNDATION

150 EAST FOOTHILL BOULEVARD, ARCADIA, CALIFORNIA

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16. Abstract

An experimental study of the interrelationships between filter operation, floc properties, and filter performance was conducted. A laboratory scale pilot plant water clarification unit including rapid mixing, flocculation, sedimentation, and filtration was used for the study. Two different clarification systems were investigated. One system had flocculation and sedimentation ahead of the filters. In the other clarification system, the pretreatment consisted of chemical addition and a brief period of rapid mixing; thus, in essence, raw water was applied directly to the filters. When operated in this latter manner, the filters were actually serving as contact flocculators. The experimental results show that an effective filtration system is combined of properly matched filter media and floc properties. Weak and non-sticky flocs, for instance, are desirable for a fine media filter; whereas, the opposite type of floc are essential for successful operation of a coarse media filter. Also, the study has demonstrated quite adequately the superior performance of a dual media filter bed as compared to that of a single media and filter.
(Continued on next page)

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16. Abstract

The technical and economic feasibility of clarification by contact flocculation are demonstrated by the study results. It is concluded that a cationic polyelectrolyte, cat-floc, is more effective than alum in accomplishing suspended solids removal under conditions of moderate headlosses. A new type of dual media filter, consisting of a very coarse, uniform coal layer on top of a shallow sand layer, performed better than either a single media sand filter or a regular well graded dual media bed. Clarification accomplished through the use of such a filter bed for contact flocculation may prove more economical than the traditional three stage clarification system, provided the raw water suspended solids content is below a maximum concentration. The dependence of filter performance on such factors as chemical dose, filtration rate, and raw water suspended solids was also examined. (EPA abstract)

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NOTES ON REPORT ARRANGEMENT

Chapters are designated by Roman numerals and Arabic numerals are used for page numbering within chapters.

Tables are numbered with Roman numerals which designate the chapter and with Arabic numerals which designate the sequence within the chapter.

Figures are assembled at the end of the respective chapters.

Appendices are assembled together at the end of the report.

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CHAPTER I

INTRODUCTION

This final report summarizes the research work conducted during the second year of a two-year project ending 30 June 1968 by the Ludwig Engineering and Science Research Foundation. During the second year the major research effort was directed toward evaluation of the factors affecting floc strength, laboratory pilot plant studies of the effect of chemical factors on conventional filtration, and laboratory pilot plant studies of contact flocculation. These investigations were oriented toward laboratory testing of accepted and acknowledged concepts rather than toward developing additional theoretical or mathematical models. This orientation recognizes that basic experimental data must be obtained and evaluated before any of the advanced filtration techniques will be put to general practical use for either water or wastewater clarification.

SUMMARY OF THE FIRST PROJECT YEAR

The first project year was devoted to determining how to measure floc strength, and to evaluating the relationship between floc strength and filtration performance. The results of the first years efforts suggest that an optimum floc strength may be found at which moderate removal efficiency and moderate headloss buildup rate are obtained. These latter factors are associated with increased filtration runs and better economy in water treatment plant operation.

Several methods for measuring floc strength were considered and two methods, the Silting Index and the median-floc-size, were studied in detail. The Silting Index method measures the decrease in filtration rate at a constant pressure as the thickness of filter cake increases. The specific resistance of the cake can be calculated and is considered as a measure of floc stiffness. In the median-floc-size method the strength of the floc is rated on the basis of the median size of floc which has been subjected to a selected shear field. The median size is determined from a size distribution obtained by subsidence rate measurements.

The two methods were evaluated using an alum floc suspension and the effect of coagulation pH and alum dose on floc strength was observed. In order to obtain design data and to evaluate the effect of floc strength on actual filter performance a laboratory pilot plant was designed and constructed. The laboratory pilot plant and the results obtained from its

operation are discussed in subsequent chapters.

RESEARCH OBJECTIVES

The general objective of the second-year investigation was to provide experimental data which would demonstrate the interrelationships between various design and operation parameters of a filtration system. This objective was realized using two different filtration systems. One system utilized flocculation and sedimentation prior to filtration. In the other system the raw water was applied to the filter after chemical addition and a very short mixing period. In this latter system, flocculation took place inside the filter; thus the filter served as a "contact flocculator."

The studies of the traditional flocculation-sedimentation-filtration water treatment system were limited to the filtration process. The factors of interest were the floc properties and their relation to the performance of filter beds composed of different types of media. In order to make the results applicable to practical filter design, all parameters and methods were selected to stimulate those encountered in the field.

The specific objectives of this phase of the study were:

(1) To determine experimentally the effect of flocculation pH on the strength and electrophoretic mobility of the flocs leaving the sedimentation tank.

(2) To examine the effects of floc properties on the performance of different filter beds.

(3) To derive the general relationships between the design, operation, and performance of filter.

Contact Flocculation

Because the contact flocculation process is in the early stages of development, the general objective of this portion of the study was to explore the technical and economical feasibility of the process. The specific objectives of this phase of the investigation were:

(1) To study the effect of alum dose on the performance of contact flocculators made of various filter media.

(2) To study the performance of contact flocculators with polyelectrolytes as filter aids.

(3) To demonstrate the interrelationships between floc properties, filter media, and filter operation parameters in contact flocculation.

(4) To compare the performance of a conventional flocculation-clarification system to contact flocculation, analyzing the merits of each under various circumstances.

PROJECT STAFF

Participants in this study from the Ludwig Engineering and Science Foundation were:

Dr. Harvey F. Ludwig, Project Director

Dr. Wonsuk Kim, Project Engineer

Dr. Yerachmiel Argaman, Project Engineer

Mr. Taghi Manbeian, Research Assistant

Mr. Stanley R. Primmer, Research Assistant

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Mr. William D. McGuire, Water Treatment and Distribution Supervisor, East Bay Municipal Utility District, Oakland, California

CHAPTER II

FLOC PROPERTIES AND FILTRATION PERFORMANCE

OBJECTIVES

This part of the research was conducted for the purpose of establishing the relationship between filter performance and floc properties for clarification systems in which filtration is preceded by flocculation and sedimentation. The properties perceived to affect most filter performance were the floc strength and the tendency of floc to adhere to the filter media (floc "stickiness"). Floc strength is indicated to be the more important of these two properties, as acknowledged by Mintz and Hudson (References II-1, and II-2). Although the importance of floc strength in filtration performance has been recognized and accepted by many investigators, experimental data defining the relationship between the two factors is relatively scarce. Thus, the objective of this phase of the study was to generate the information necessary to develop the relationship by comparing filtration performance with floc strength data obtained independently of the filter operation. It is logical to assume that the effect of floc strength on filtration performance will be system dependent. The principal causes of this system dependency are the effects of the hydraulic through-put rate and the geometric properties of the media. Therefore, it was not expected to find a single optimum floc strength applicable to all cases and it was appropriate to examine several different filter media. A major consideration in the selection of experimental conditions and techniques was the desire to obtain information which would be readily applicable to full scale filtration units. Thus, the filter media, the pretreatment methods, the raw water quality, and the chemicals used were selected to be as comparable as possible with those experienced or used in field applications.

APPARATUS

The experimental work was carried out in the pilot plant shown in Figures II-1 and II-2. The plant consisted of a 1.0 gpm raw water supply whose quality could be controlled, flocculation and sedimentation chambers, and four 4.0-inch inside diameter filter columns each containing a different filter media. The composition of the filter media used for this phase of the study is shown in Figure II-3. An effluent flow rate control device was used to maintain a constant flow rate throughout each run. A schematic diagram of

the flow rate control device is presented in Figure II-4. Provision was made for sampling the water at different points in the filter columns and for measuring the piezometric heads at various depths. The sampling ports consisted of syringe needles which were introduced into the bed through rubber stoppers mounted in the filter wall. The design and operating characteristics of the various elements of the pilot plant were as follows:

(1)	Flash Mixing	
	Residence time*	0.32 min
	Mixing speed	120 rpm
	Paddle size	1.5 x 2.0 in
(2)	Flocculation Reactor	
	Overall residence time*	23.1 min
	Number of compartments	3
	Estimated root mean square (rms) velocity gradient	39 sec^{-1}
(3)	Settling Tank	
	Residence time*	45 min
	Surface loading	240 gpd/sq ft
(4)	Filters	
	Surface loading	3.0 gpm/sq ft
	Rising velocity in backwash at 50 percent bed expansion	
	Filter No. 1	43 in/min
	Filter No. 2	53 in/min
	Filter No. 3	28 in/min
	Filter No. 4	28 in/min

* Theoretical hydraulic detention period at 1.0 gpm

EXPERIMENTAL PROCEDURES AND DATA COLLECTION

A synthetic raw water containing 6 mg/l clay was flocculated with 20 mg/l alum, settled and applied to the filters through the constant head distribution box. The flow rates were adjusted to predetermined values using calibrated rotameters. Influent turbidity and particle mobility analyses were performed on samples of floc suspension obtained from the distribution box. Turbidity measurements were made using a Hach Model 1806 turbidimeter. A standard Briggs Cell was used to determine particle mobility. The floc strength was estimated by comparing the median size of the unsheared flocs to that of a floc suspension which had been fragmented by subjecting it to a mean shearing stress of 11.5 $dyne/cm^2$ in a capillary tube. The turbidity of

the water at various depths in the filters was measured periodically and recorded along with the piezometric heads at the same depths. Prior to each run the filters were preconditioned by passing the floc suspension through them for 12 hrs followed by backwashing. The preconditioning step was recommended by Fox and Cleasby (Reference II-3) to assure reproducibility of results.

PREPARATION OF RAW WATER

The chemical and physical properties of the synthetic raw water used in this study were adjusted to simulate a local surface source. The primary criterian for similarity was the coagulation behavior at varying pH values but with a constant alum dose of 20 mg/1. Figure II-5 shows the results of standard jar tests performed on five different suspensions in local tap water including the raw water from the Sobrante Filter Plant which was the water used as the basis of simulation. Performance of the pilot plant was compared to that of the Sobrante filtration plant of the East Bay Municipal Utility District. Of the four different suspensions examined, the one showing the closest similarity to the Sobrante water is represented by curve "D" of Figure II-5. Based on these results, this synthetic raw water was selected for use in the study and is designated as the raw water. This suspension was prepared by adding the following chemicals to the local tap water:

Kaolinite clay	2.0 mg/1
Bentonite clay	4.0 mg/1
$NaHCO_3$	1.12 meq/1
Na_2SO_4	0.8 meq/1
HCL	0.12 meq/1

The quality of tap water is sampled weekly at a site near the laboratory (Oakland Tap 9) by the local purveyor (East Bay Municipal Utility District). The characteristics of tap water during the first quarter of 1968 were (average of weekly analyses): alkalinity , 77 mg/1; chlorides, 9 mg/1; hardness, 100 mg/1; turbidity, 0.2 JTU; pH, 8.0. Periodic analyses of the synthetic water were conducted to assure its constant quality. The various chemicals were added to the local tap water by means of a sigma pump. Floc properties were varied by adjusting the pH of the raw water through the addition of either acid or base.

EXPERIMENTAL RESULTS

Effect of pH on Floc Properties

The pH of the water in the flocculation chamber affected the concentration and the physical characteristics of the floc suspension entering the filter columns. This is illustrated in Figure II-6. The upper curve of this figure shows the variation of the sedimentation tank effluent turbidity as a function of pH. These variations reflect the effect of pH on the flocculation process and the settleability of the flocs. The best flocculation-sedimentation performance was obtained at pH values of from 6.0 to 7.0 as indicated by the low settled water turbidities at these pH values. Comparison of the upper and middle curves of Figure II-6 shows that the best flocculation was associated with low mobility i.e., low zeta potential. The pH values at which the best flocculation performance was obtained were very close to the iso-electric pH of 6.5.

The influence of pH on floc strength is illustrated in the bottom curve of Figure II-6. The median size of the unsheared floc increased rapidly as the pH was changed from 5 to 6, but increased at a significantly reduced rate as the pH was further increased to 7. This indicates that the floc size is both large and nearly constant at pH values close to the iso-electric point. A very distinct difference between the size of sheared and unsheared flocs was observed at pH 7. When compared to the insignificant size difference between sheared and unsheared floc at pH 5 and 6, the size difference between these two types of floc at pH 7 indicates that the floc formed in the region of the iso-electric point was very weak. The electrical charge on the particle surface is one of the factors which may effect the attachment of particles to the filter media. Considering filter media particles have a negative charge, positively charged particles might be expected to be removed more effectively than negatively charged ones. The change in particle mobility as a function of pH (Figure II-6) indicates a shift from positive to negative charge as the pH was increased above 6.5 Thus, better filter performance might be expected at pH values below 6.5.

Filter Performance and Floc Properties

The duration of a filter run (time between backwashings) is an accepted measure of filter performance. A filter is backwashed whenever the total headloss or the effluent turbidity exceed a predetermined value. For the present investigation, a headloss of six feet and effluent turbidity of 0.2 Jackson Turbidity Units (JTU) were selected as the limiting values. The terms T_B and T_H represent the time required to experience floc breakthrough and limiting headloss respectively, as determined by data extrapolation. Thus,

these values do not necessarily represent physically observed data. The filter performance, T was determined by the smaller of the two time intervals. The experimental results expressed in terms of T_B and T_H are summarized in Table II-1. The variations in the duration of the filter runs reflect the overall performance of the clarification system which includes flocculation, sedimentation, and filtration. The effects of pH on the performance of the filters can be evaluated by comparing filter runs in which the floc concentration entering the filters was the same. Such a comparison is presented in Figure II-7.

The relationship between the duration of a filter run and pH is determined by both floc strength and the ability of the floc to adhere to the media. The former was estimated by comparing the size of sheared and unsheared floc and the latter was evaluated on a relative scale using particle mobility. For any given situation either floc strength or floc adherence will be the dominant factor in determining the duration of a filter run. Based on the data in Table II-1 it is apparent that in almost all cases floc breakthrough controlled the duration of the filter runs. Either inadequate floc strength or poor floc adherence cause floc breakthrough; thus, a decrease in either of these properties will result in shorter filter runs. The influence of these properties on filter performance is illustrated in Figure II-7. For an influent turbidity of 2.0 JTU Filters 2, 3, and 4 gave maximum performance at pH 5, but with Filter 1 no optimum pH was observed. The data presented in Figure II-7, for the influent turbidity concentrations of both 1.0 and 2.0 JTU, strongly suggest that the removal mechanism in effect in all the filters was surface straining at the top of the sand media. Thus, the factor controlling the duration of the filter runs would have been the floc strength. The possible exception to the preceding is Filter 1. For this filter the size of the floc appears to have been sufficient to compensate for the reduced floc strength at the higher pH values. This compensatory effect would have been nullified in Filters 3 and 4, which had the same sand as in Filter 1, because the larger floc would have been removed in the coal media section of the filters.

Filter Media, Floc Properties, and Design Balance

A well functioning filtration system is realized when a filter bed has been designed to match the physical properties of the suspended matter to be filtered. Thus, because the properties of suspended matter are not universally constant there is no single filter design which can be designated as

TABLE II-1

EFFECT OF pH ON FILTER PERFORMANCE

Run No.	pH	Filter No. 1		Filter No. 2		Filter No. 3		Filter No. 4	
		T_B(1)	T_H(2)	T_B(1)	T_H(2)	T_B(1)	T_H(2)	T_B(1)	T_H(2)
1	4	21.5	51.6	14.5	72.4	23.0	62.4	22.2	124.0
2	5	16.2	16.9	24.7	24.0	28.7	29.0	25.9	63.0
3	6	27.0	26.5	20.1	46.8	34.1	41.5	37.1	74.5
4	7	30	31.6	14.8	52.0	21.5	47.0	23.1	101.0
5	8	11.1	12.7	8.2	22.0	17.0	17.2	23.0	46.0
6	9	0.0	-	0.0	-	0.0	-	0.0	-

NOTES: (1) T_B: Time to reach floc breakthrough, hrs.

(2) T_H: Time to reach six feet headloss, hrs.

optimum for all circumstances. The ideal filter is that which retains a significant portion of the influent solids while exhibiting a minimal rate of increase in headloss. In terms of filtration times this means high values of both T_B and T_H. As is apparent in Figure II-8, T_B and T_H depend on floc properties as well as filter design parameters. A change in any of these parameters which causes T_B to change in one direction will generally result in an oppostie change in T_H. Balanced filter design is attained when T_B and T_H are essentially equal. However, a balanced filter design does not necessarily insure long filter runs or even filter runs which are acceptable. The time to reach floc breakthrough (T_B) will usually increase with increasing floc strength, floc stickiness, and filter depth; and decrease with increasing media size. The reverse of the preceding relationships prevails when the dependent variable is the time to reach the limiting headloss (T_H).

Figure II-8 shows T_H plotted against T_B. The dashed line in the several graphs represents the "line of balanced design". If the points fall along this line then the hydraulic and media characteristics of the filter are suitably matched with the floc properties. Points which fall above the line of balanced design represent situations in which floc breakthrough occurred before the limiting headloss was reached; and points which fall below the line indicate filtration conditions in which headloss was the time limiting factor. These relationships are system dependent in that whether or not the design is balanced depends in part on the selected limiting headloss and turbidity.

Filter No. 1 exhibited balanced design at all pH values except at pH 4. Filter No. 4 experienced floc breakthrough before attaining maximum headloss at all pH values. At pH 4, 6, and 7 the filtration runs of Filters No. 2 and 3 were terminated by floc breakthrough. Filter No. 3 exhibited balanced design at pH 5 and 8. Filter No. 2 reflected balanced design characteristics at pH 5; however, at pH 8 floc breakthrough occurred before the limiting headloss was realized.

The media of Filter No. 1 is typical of that used in many water filtration facilities. Because balanced design was realized under several different conditions it appears that floc strength, floc size and media penetration are compensatory within the headloss constraint selected. At pH 4 the floc was too small and possibly too weak to be retained on the surface of the media under an increasing hydraulic gradient. The effect of

media size on balanced design is illustrated by the difference between the behavior of Filter No. 1 and Filter No. 2. Except for pH 5, Filter No. 2 experienced floc breakthrough before Filter No. 1 and it also exhibited a much lower rate of headloss buildup than Filter No. 1. At pH 5 the flocs were of sufficient size and strength to be retained primarily on the surface until the limiting headloss had been reached. It is also possible that during this particular filtration run the floc which was sheared was still of a size to be retained in the body of the media.

Filter No. 3 had the same size sand media as Filter No. 1. The depth of sand media was two inches less in Filter No. 3 than in Filter No. 1; the top two inches of Filter No. 3 was occupied by a coal layer. The apparent effect of this coal layer was to decrease headloss buildup and to increase the time to floc breakthrough as compared to those characteristics of Filter No. 1; the exception being the run at pH 7 with very weak floc of slightly negative mobility. The performance characteristics of Filter No. 3 are a result of the larger floc being removed by the coarse anthracite layer resulting in a reduced load to the top of the sand media. Filter No. 4, composed of 18 inches of anthracite and six inches of sand, exhibited lower rates of headloss buildup and slightly longer times to floc breakthrough than Filter No. 3. The performance of Filter No. 4 appears to be an extension of those observed for Filter No. 3; this being a result of the increased depth of anthracite in the former as compared to that in the latter.

The distribution of the floc deposits through the depth of a filter bed is implied by headloss profiles such as those shown in Figure II-9. The figure shows the net rate of headloss per unit filter length due to partial clogging of the pores by deposited flocs for each of the four filters at two different pH values. Values of net headloss per unit length were calculated by subtracting the clean water headloss from the total headloss at terminal conditions. For the runs at both pH 6 and 7 the concentrations of flocs in the filter influent were the same, but the floc properties were different. At pH 6 the flocs were stronger and were positively charged; hence, they were less likely to penetrate deep into the filters. Indeed, the data of Figure II-9 show that the floc deposition in the upper portion of the filter beds was more pronounced at pH 6 and pH 7. At both pH values floc accumulation of the top layers was greatest in Filter No. 1, the single media fine sand filter. Filter No. 2 with the coarser media showed a better utilization of the entire filter bed, although the bulk of the floc

accumulation still occurred in the top four inches of the bed. A much better utilization of the total media was obtained in Filter No. 4, the dual media filter. The addition of a 2-in layer of coal in Filter No. 3 resulted in maximum floc accumulation at about 4.0-in below the surface. It should be noted that although the sand layer in this filter was of the same characteristics as in Filter No. 1, the maximum headloss was much smaller. This variance is probably due to the formation of a mixed media zone at the coal-sand interface of Filter No. 3, eliminating the zone of very fine sand which existed in Filter No. 1.

SUMMARY

The results of the experimental work have demonstrated the manifestations of some of the interrelationships between filter design, filter performance, and floc properties. The only general conclusion which can be drawn from the results is that all design and performance parameters are interrelated and optimization of filter operation requires consideration of all factors simultaneously and a clear delineation of the objective function. Neither a particular filter media nor a particular floc can be considered as having optimum characteristics in an abstract context.

If the flow rate and the raw water turbidity are fixed the length of filter run becomes a relative measure of filter performance. A filter run is terminated whenever the headloss or the effluent turbidity exceed a predetermined limit. The experiments have shown that early floc breakthrough is associated with weak or non-adhering flocs and coarse media. Rapid buildup of headloss is associated with strong or sticky flocs and fine media. The proper design and operation of a filter involves appropriate combination of floc properties, filter media, flow rate, and limiting values of turbidity and headloss.

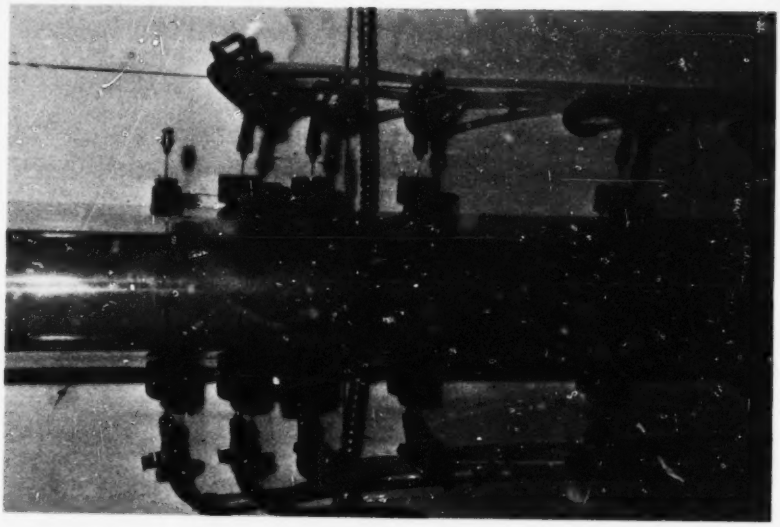
A filtration process is described as having a balanced design if the limiting headloss and floc breakthrough occur at the same time. In most cases balanced conditions can be approached by proper operation of a given filter. The "typical" sand filter of the present study showed a balanced operation over a wide variety of conditions. This is attributed to compensation among the floc properties, hydraulic conditions, and filter media. Other filters, although not operating at balanced conditions, experienced filter runs of longer duration than Filter No. 1. The longest filter run was obtained with the dual media Filter No. 4 at pH 6. This run lasted for 37 hrs as compared to a maximum of 30 hrs for Filter No. 1. It is apparent

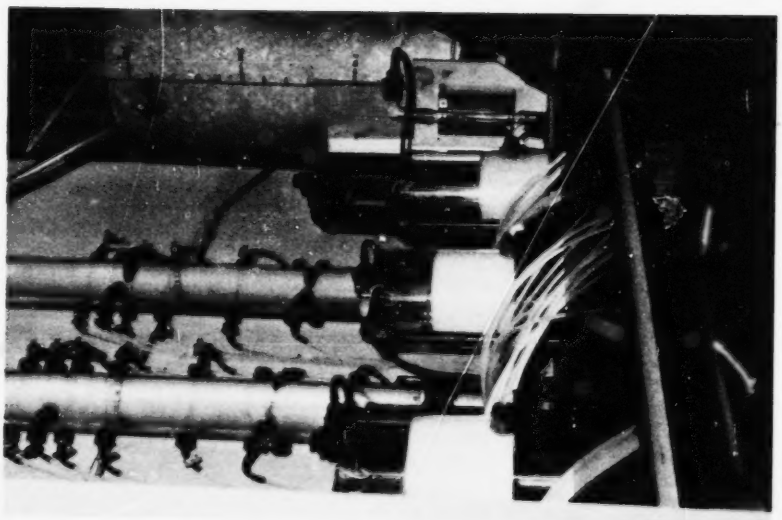
that design balance in itself does not guarantee long filter runs.

The experimental results of this phase of the study have demonstrated the basic weaknesses and poor design involved in the traditional filtration systems. Flocs entering the filter are usually collected at the top layers and the larger portion of the bed remains essentially unused. Those flocs which do penetrate deeper into the bed are usually small and exhibit insufficient adherence to be removed; thus, they appear in the filter effluent. In the common fine-to-coarse filter, the probability of the floc escaping the filter increases with depth. That is, the deeper a particle penetrates into the filter bed the less is the chance that it will be removed in the next increment of depth. In a dual media filter with a fine-to-coarse graduation of each media, the same filter inadequacy exists, although to a lesser extent. The larger pore sizes are not the only cause of reduced removal efficiency in the deeper layers of the filter media. Of equal importance in effecting reduced removal is a reduced adherence tendency which is typical of aged flocs. The floc which reaches a filter is relatively well aged and exhibits a degree of stability which it did not possess when it was initially formed. Thus, it has very little tendency to agglomerate with other flocs or to attach strongly to filter media with which it might come in contact.

In essence, the major operational inadequacies of present standard filtration systems are the inefficient use of the media and the unsuitable condition of the floc reaching the filter. Therefore, in order to recognize improvements in the filtration process it is necessary to devise a filtration system which promotes more efficient use of the media and provides both a better quality floc and better control over floc formation. Because the contact flocculation or depth filtration process appears to have the facility to accomplish the preceding objectives, the second phase of this study was devoted to considering this process.

PILOT PLANT





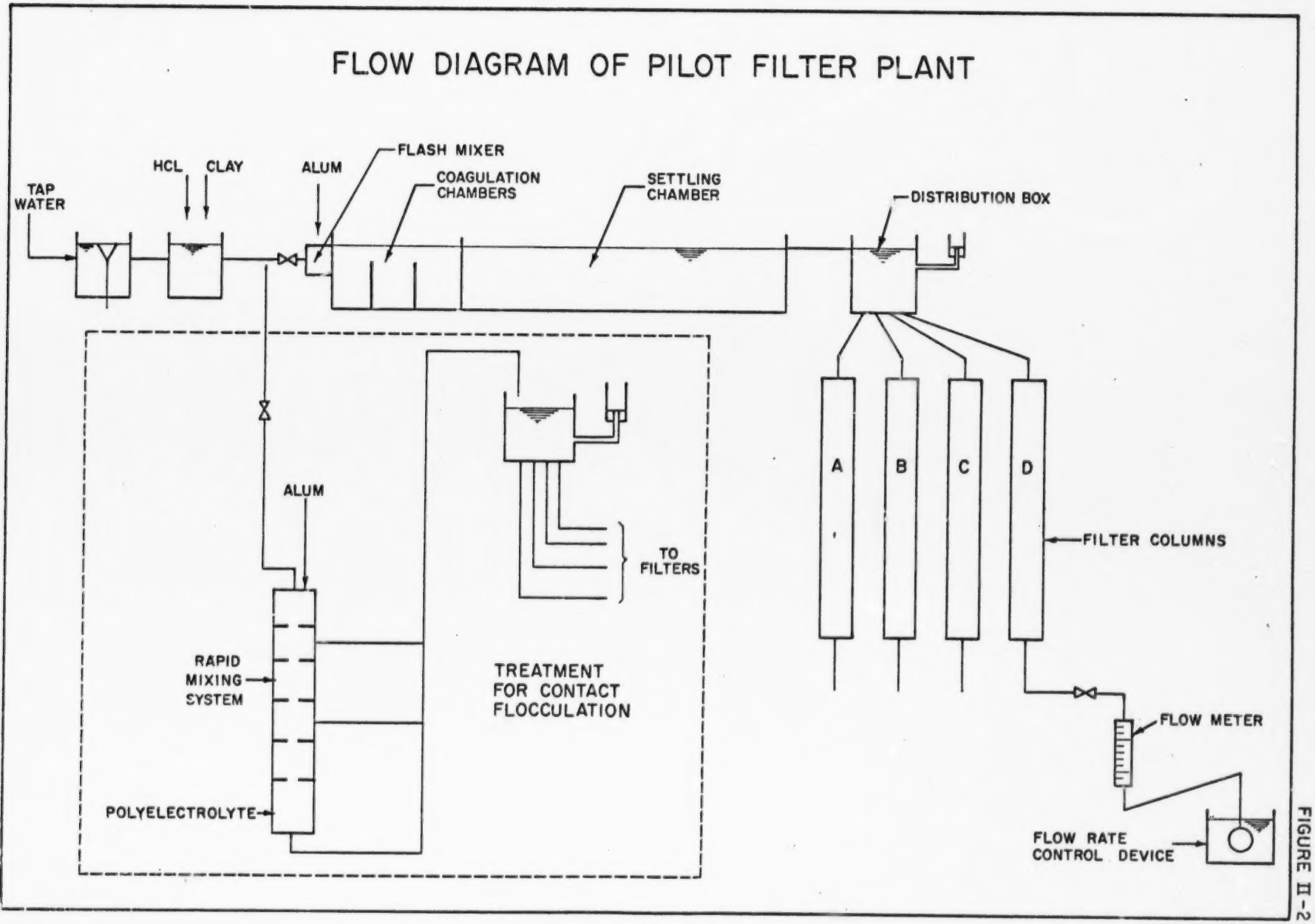

FLOW DIAGRAM OF PILOT FILTER PLANT
TAP WATER
HCL
CLAY
ALUM
FLASH MIXER
COAGULATION CHAMBERS
SETTLING CHAMBER
DISTRIBUTION BOX
A
B
C
D
FILTER COLUMNS
ALUM
TO FILTERS
RAPID MIXING SYSTEM
POLYELECTROLYTE
TREATMENT FOR CONTACT FLOCCULATION
FLOW METER
FLOW RATE CONTROL DEVICE

FIGURE II-2

COMPOSITION OF FILTER MEDIA

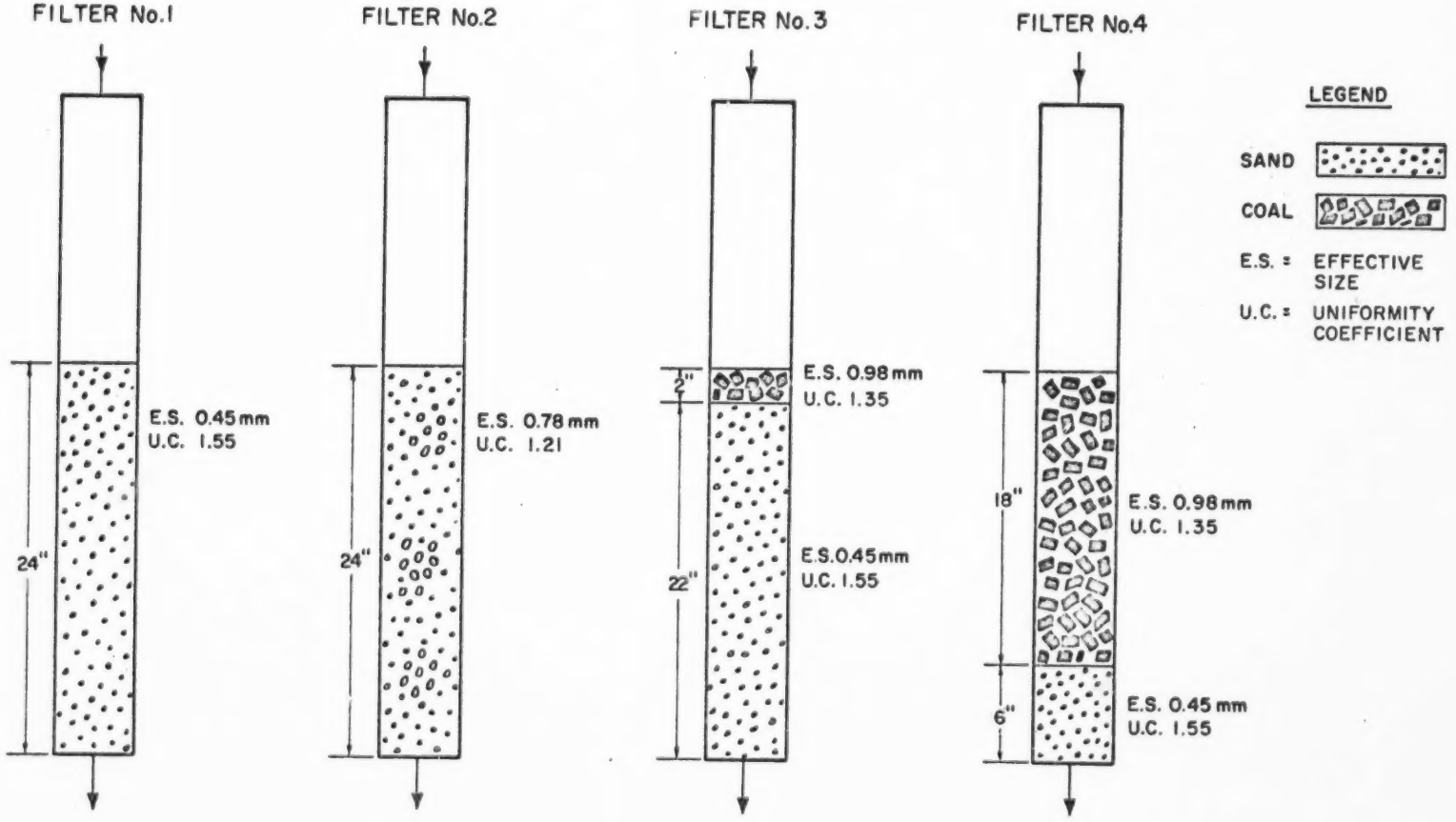
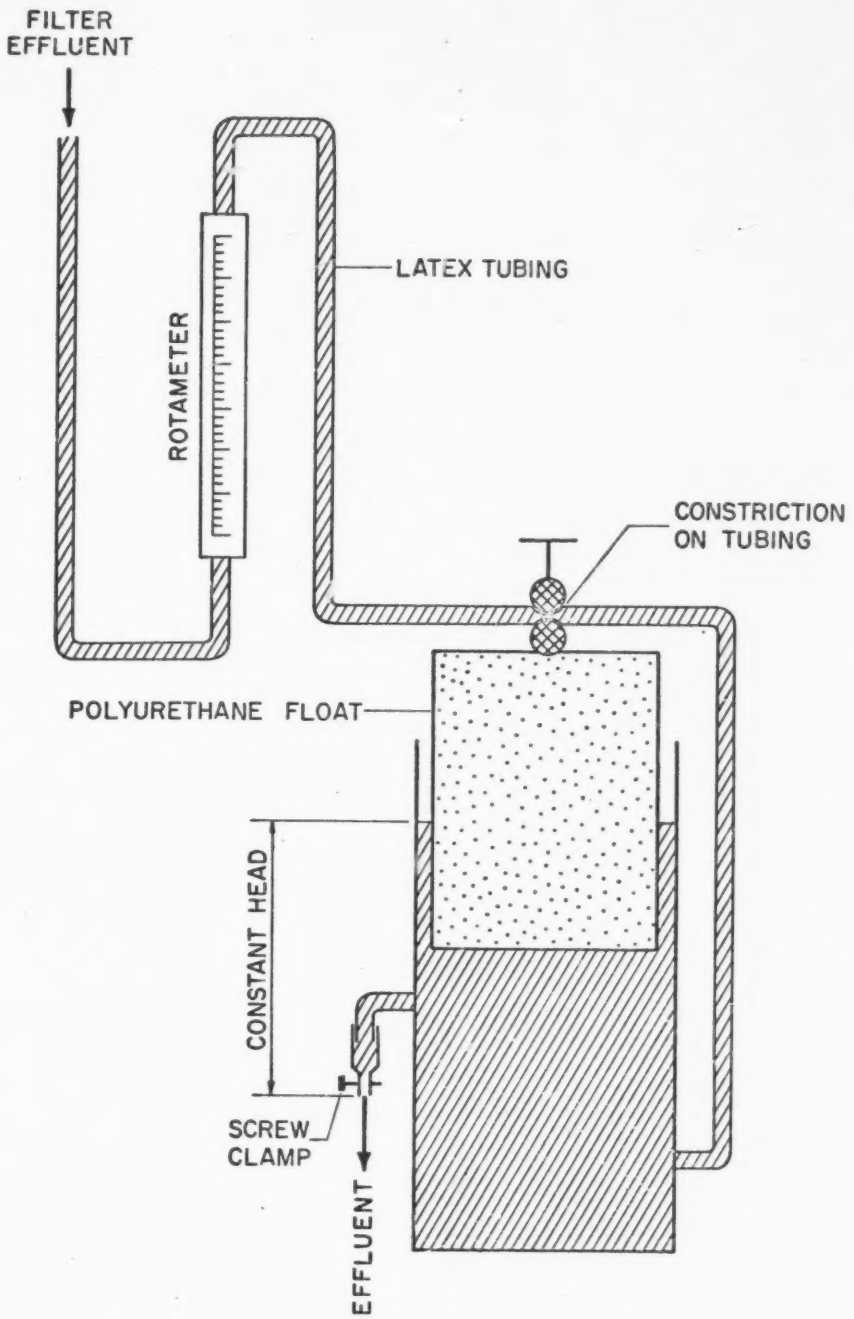
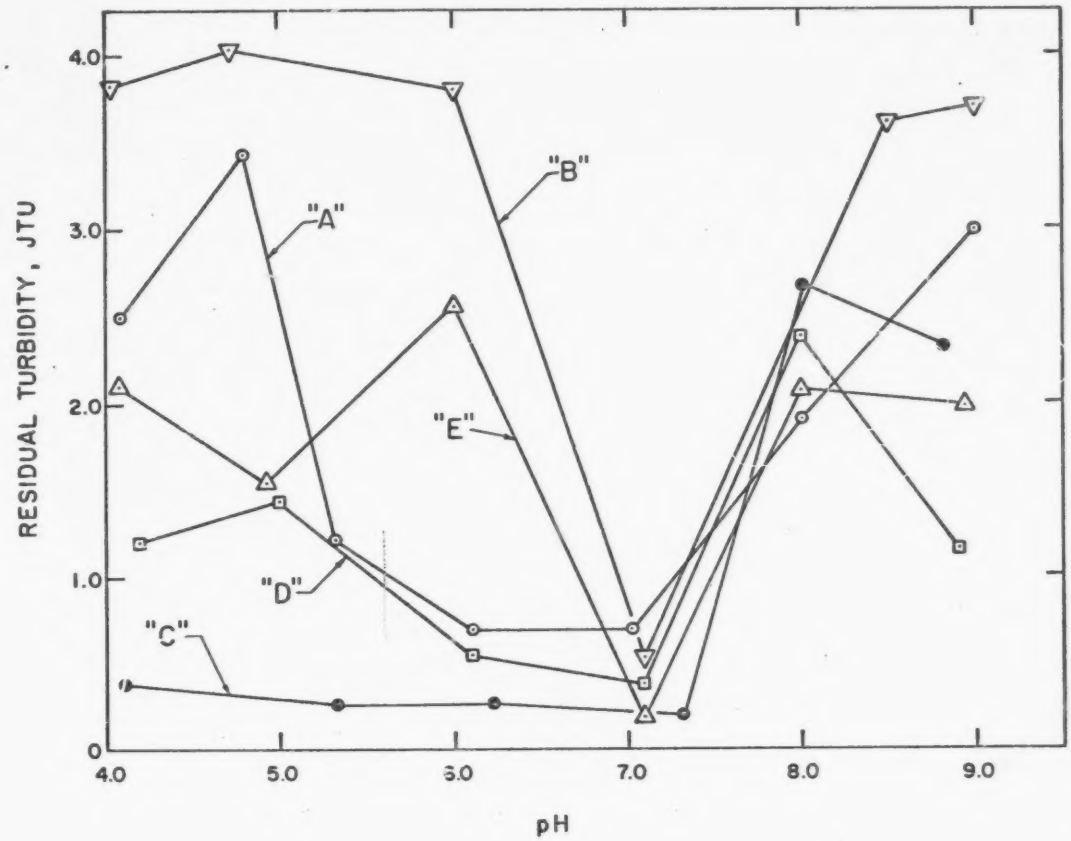


FIGURE II-3

FLOW RATE CONTROL DEVICE



JAR TESTS TO SIMULATE LOCAL SURFACE SUPPLY



A. Raw Water from Sobrante Filter Plant

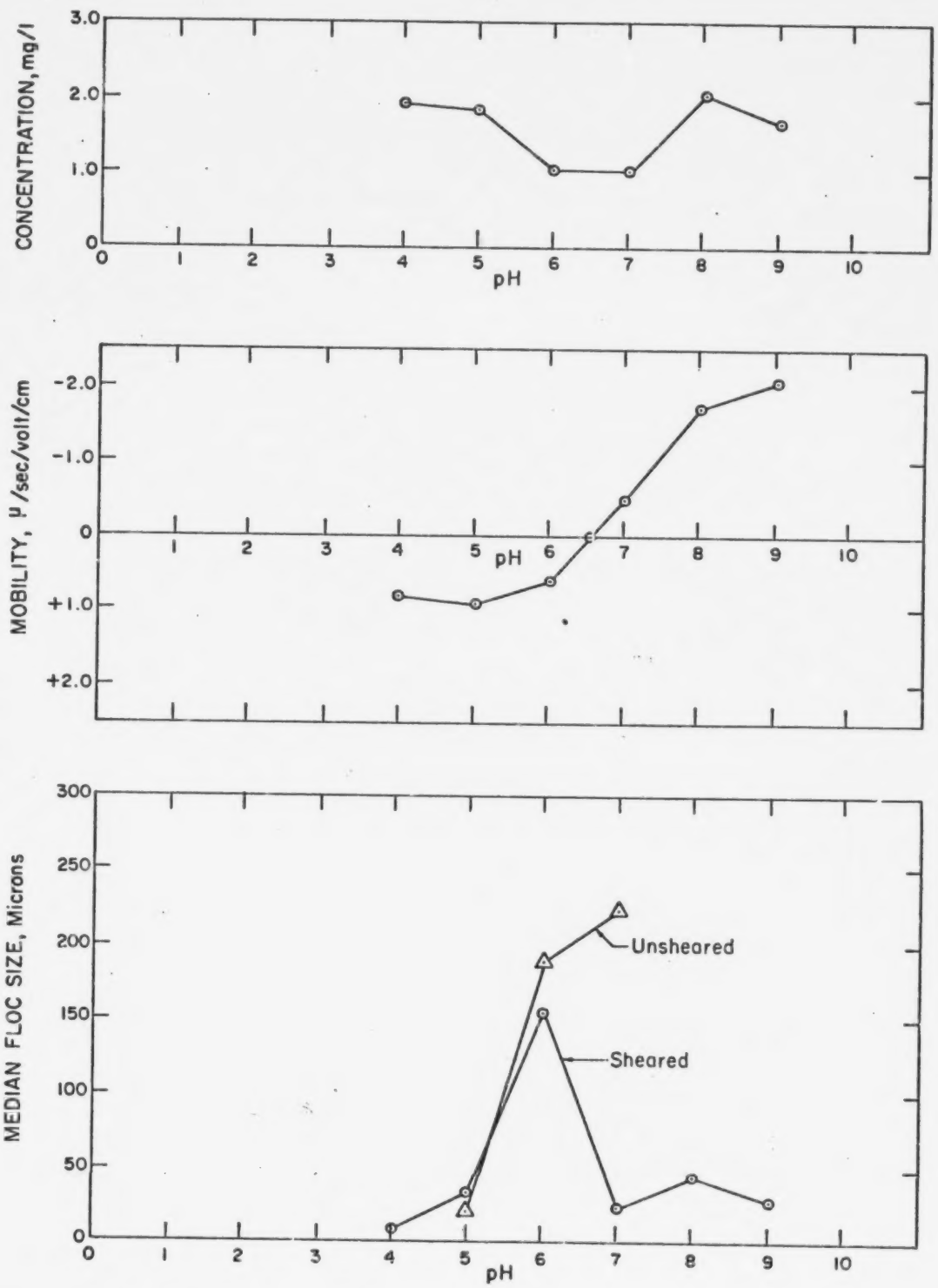
B. Kaolinite Suspension, 3.0 mg/l

C. Bentonite Suspension, 2.0 mg/l

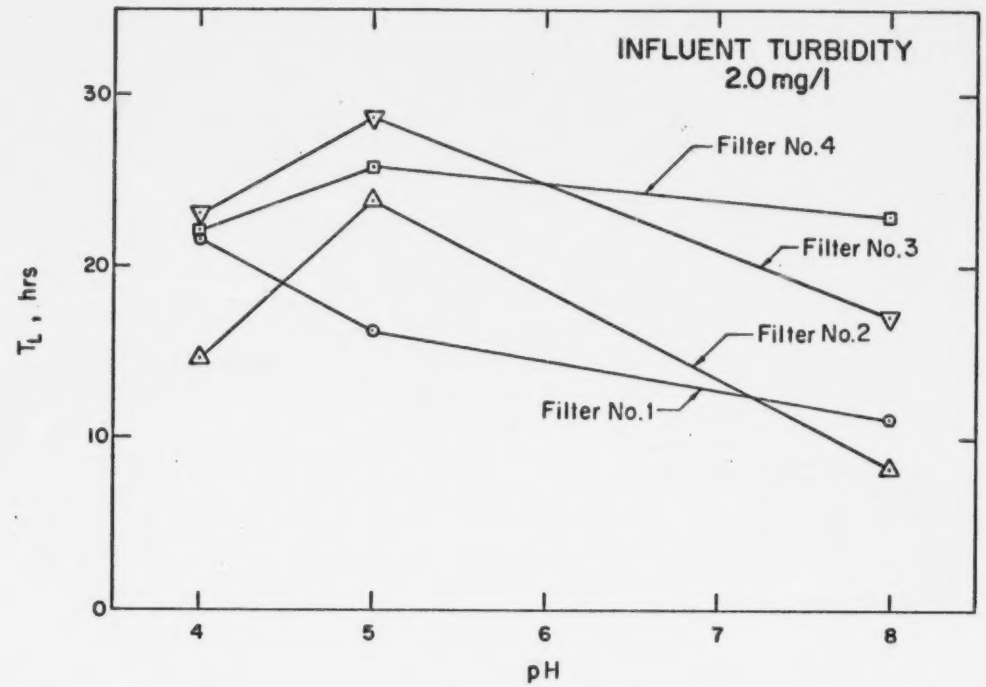
D. Kaolinite 2.0 mg/l + Bentonite 4.0 mg/l
$Na\,HCO_3$ 1.12 meq/l
$Na_2\,SO_4$ 0.8 meq/l
HCL 0.12 meq/l

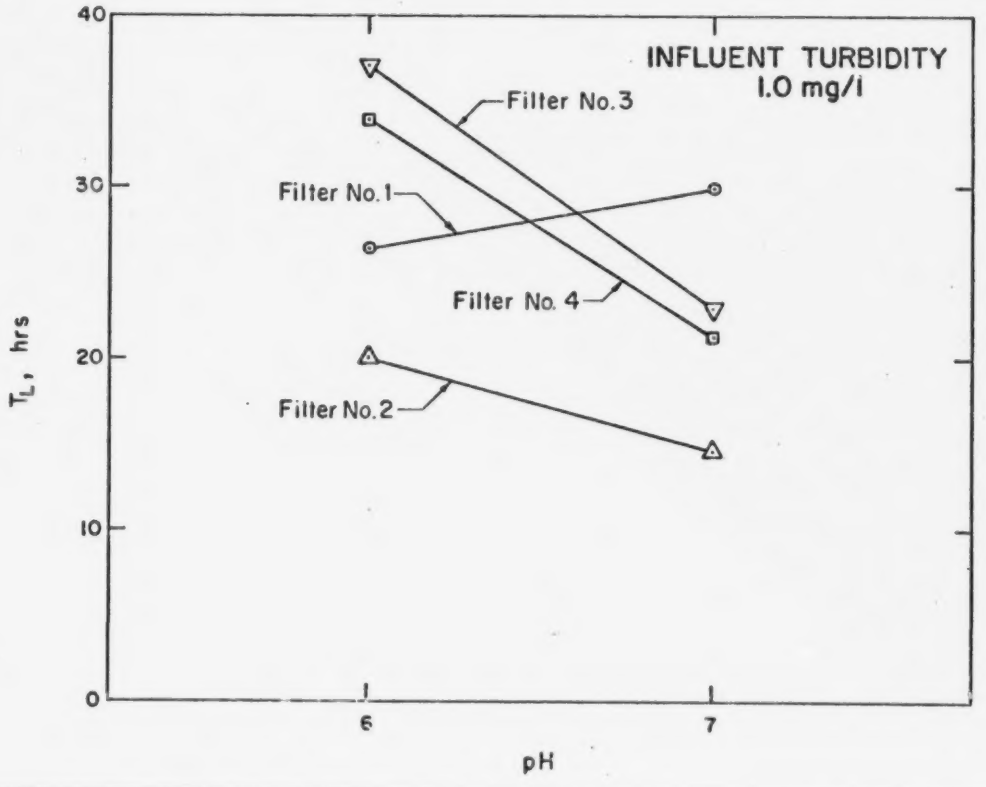
E. Same as D without $Na_2\,SO_4$

VARIATION OF FLOC PROPERTIES WITH pH

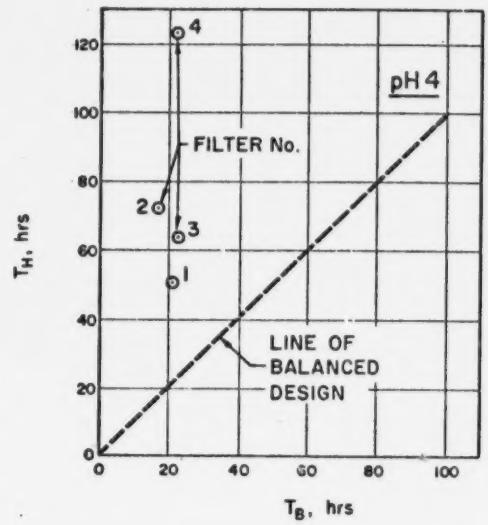


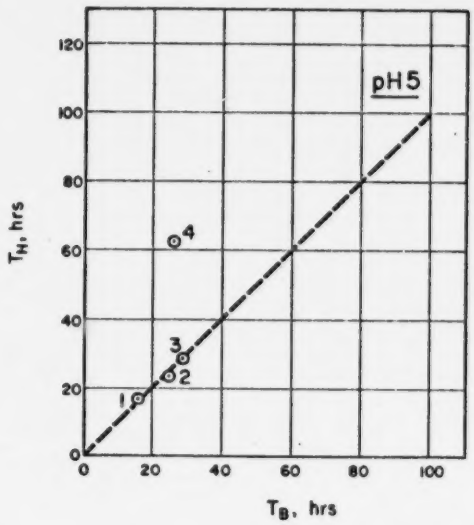
EFFECT OF pH ON FILTER PERFORMANCE

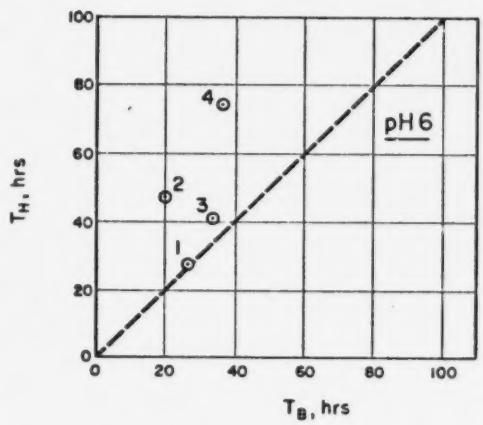


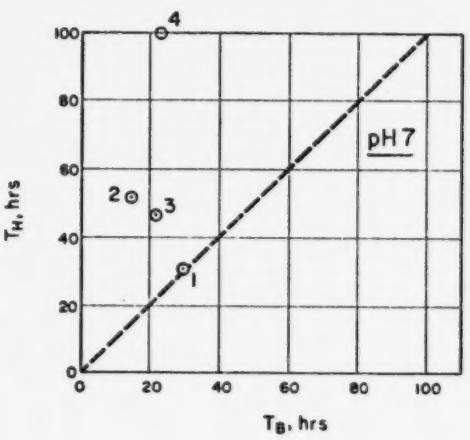


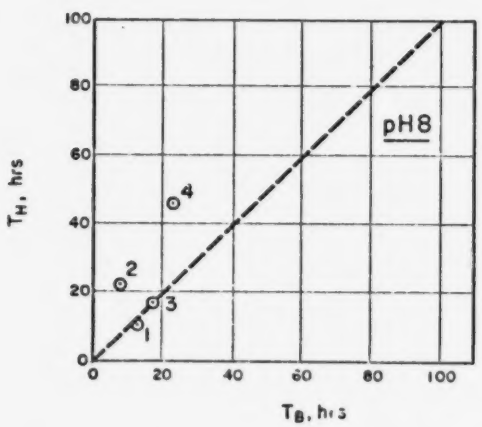
FILTER MEDIA AND DESIGN BALANCE*

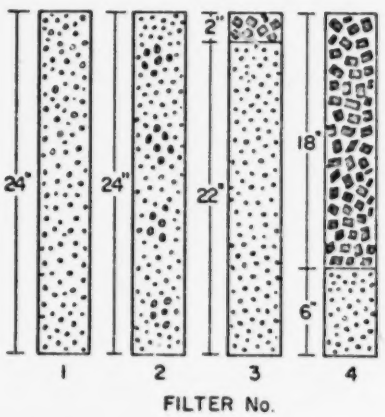










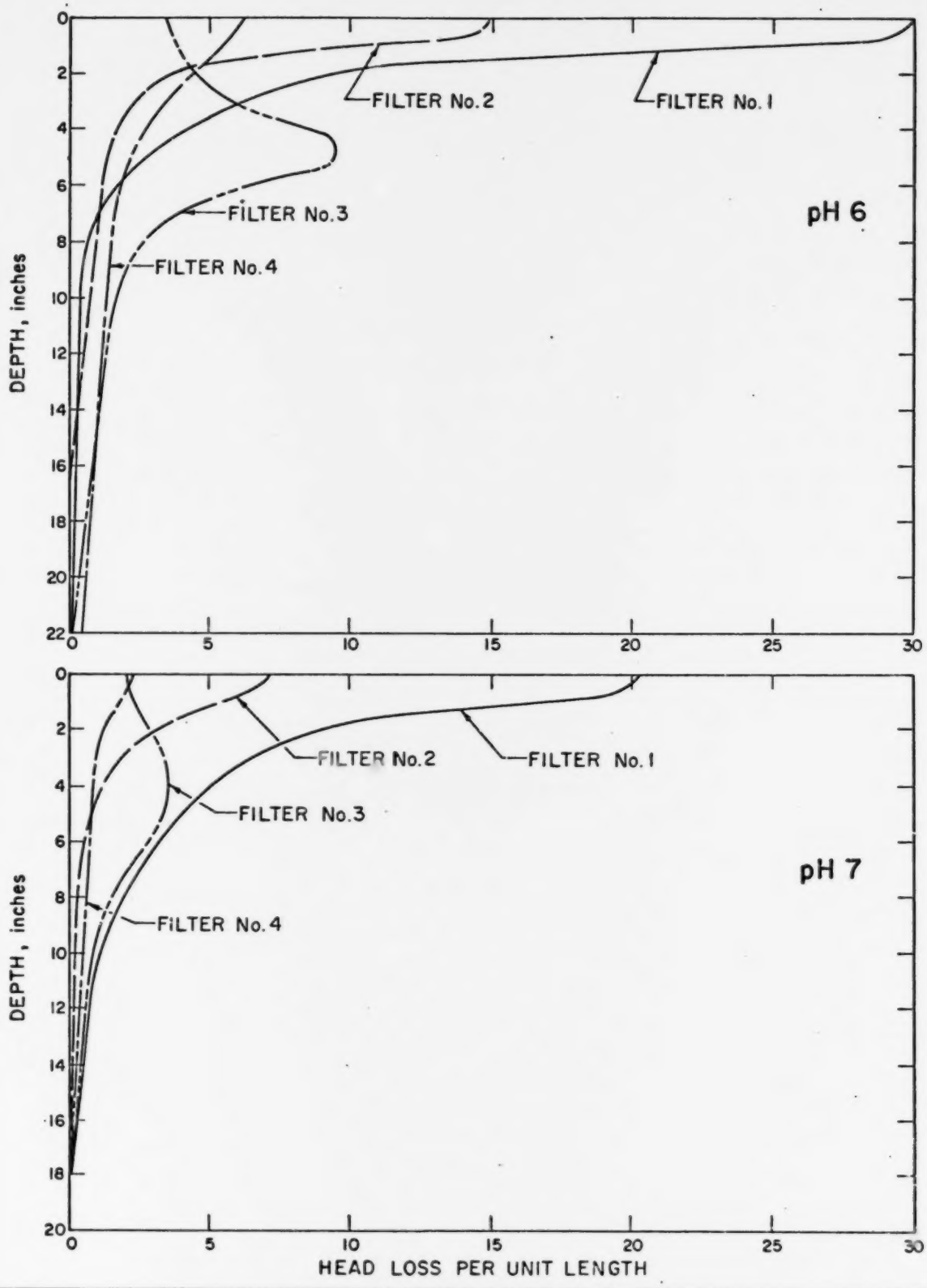


NOTE: ⊙4 Indicates Filter Numbers

* Based on limiting headloss of 6.0 ft. and filter effluent turbidity of 0.2 JTU.

FIGURE II-9

HEAD LOSS PROFILES AT TERMINAL CONDITIONS



ENGINEERING - SCIENCE, INC.

CHAPTER III

CONTACT FLOCCULATION

THEORETICAL AND TECHNOLOGICAL BACKGROUND

In the water industry water clarification has been effected by the traditional three step system consisting of coagulation-flocculation and two separation processes, sedimentation and filtration. In the coagulation-flocculation process, the suspended colloidal particles are aggregated to form larger flocs which are amenable to separation from the liquid phase by one or the other of the separation processes. The first separation process, sedimentation, accomplishes removal of those flocs which have a sufficient subsidence velocity to be eliminated from the process overflow. The second separation process, filtration, removes those particles in the sedimentation process overflow. The quantity of solids, which are available for removal by the filtration process, is relatively small. Thus, in the traditional rapid sand filter plant, filtration is, in essence, a polishing process.

The conventional single media sand filter is designed and operated in such a way that during backwash the sand grades hydraulically with the finest particles at the top of the bed. As a result, the top layer of the fine media effectively removes most of the incoming flocs allowing very few to penetrate to the deeper media layers. The accumulation of flocs at the top layer of the bed causes a rapid buildup of headloss due to clogging of the pores. When the headloss or effluent turbidity exceeds a predetermined value, the filter has to be backwashed. Frequent washings or short filter runs impair the system's economy.

With fine-to-coarse filtration, only a small portion of the filter bed is utilized as almost no floc penetration takes place. Some improvement of the system's performance may be achieved by increasing the hydraulic loading on the filter. The higher fluid velocities and the associated shear stresses cause some floc penetration which results in a better utilization of the filter bed. Still, most suspended matter is retained by the upper part of the bed. The only way to eliminate this is to use a filter bed in which the flow is from the coarse to the fine media. With a single media filter this can be accomplished only by upflow filtration. Because this particular flow pattern poses some technical problems, an alternative way to obtain coarse-to-fine filtration is through the use of multi-media filter beds.

With the proper combination of media types consisting of fine particles of high density and coarse particles of low density, a filter bed in which the number of particles per unit volume increase with depth can be constructed and maintained. Such a filter is considerably more efficient than the traditional fine-to-coarse filter because its entire volume can be utilized for collecting the suspended material without excessive headloss within any particular layer.

Because of the considerably higher capacity of the coarse-to-fine filter to retain suspended material, it is logical to examine the merits of eliminating the traditional flocculation-sedimentation units to allow the entire suspended solids mass to be removed by the filter. The technical feasibility of this method has been established by several investigators (References III-1 and III-2) as well as by the experimental studies of the present project. Although the traditional flocculation reactor is eliminated from the system, the flocculation process still takes place within the filter bed, hence the name "contact flocculation".

In the traditional clarification system the bulk of the suspended matter is removed in the settling tank. In contact flocculation process the settling unit is eliminated and the entire suspended solids load is applied to the filter. In order to avoid rapid clogging of the filter bed, it is desirable to minimize the floc volume. For this reason contact flocculation is particularly promising in combination with polymer flocculation. The commonly used iron and aluminum salts may account for a large fraction of the volume of sludge produced in settling tanks treating coagulated water. For example, raw water containing less than 10 mg/l suspended solids, which are relatively common, when treated with 25 mg/l alum will produce sludge of which more than 50 percent may be a precipitate of aluminum hydroxide. When polyelectrolytes are used as the sole flocculants, they are applied in small doses, usually in the 1.0 mg/l range, and the sludge is composed almost entirely of solids which originated in the raw water.

In the contact flocculation process, the filter bed serves as the flocculation reactor and provides for floc storage. In order that flocculation can take place in the filter bed, electrolytes or polyelectrolytes must be added to the influent suspension just ahead of the filter. If flocculants are not added, the colloidal particles, which are too small

to be strained out and are not sufficiently adhesive, will undergo little reduction in concentration as they pass through the filter. However, if the suspension is properly conditioned, the separation process can take place in the filter.

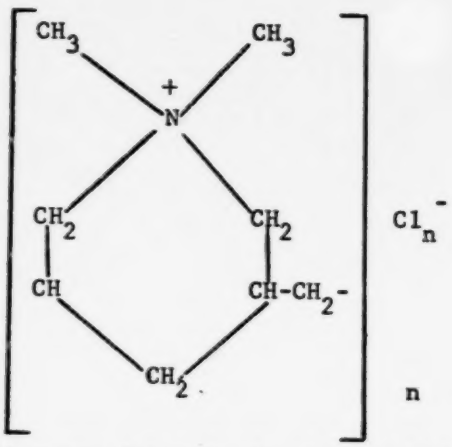
The mechanism by which the suspended solids are separated from the liquid phase and deposited on the filter media has been the subject of numerous studies. In recent papers, O'Melia and Stumm (Reference III-3) have presented a summary of existing theories of filtration along with their own proposed model. According to these authors the removal process must be considered as a combination of particle transport and attachment. Particle transport is a physical process governed by the hydrodynamics of the flow within the pores of the media. The attachment step involves chemical and electrical interactions of a very complex nature. The removal efficiency depends on the extent to which each of these steps is successful. In contact flocculation, flocs are formed inside the filter; hence, during the transport step some flocculation may take place. There is ample experimental evidence to show that large flocs are formed within the filter pores. Whether these are formed in the liquid phase prior to their deposition or buildup on the media after attachment is not clear. However, it appears that the bulk of the floc growth occurs on the media, although some aggregation does take place prior to deposition.

The present study of contact flocculation was limited to exploring the practical feasibility of the process, and to investigating the interrelationships between some of the operating and performance parameters. No attempts were made to resolve fundamental questions or to develop new or additional theoretical models of contact flocculation.

APPARATUS AND EXPERIMENTAL PROCEDURES

The apparatus used for the contact flocculation studies is described in Figure II-2 with the alternative pretreatment system. The water supply was the same as in the previously described experiments. Two chemicals were employed in different combinations. Alum was added at the top of the flash mixer which provided 30 seconds of rapid mixing with the clay containing water. A cationic polyelectrolyte was used as filter aid and was added at the lower portion of the flash mixer such that 7.5 seconds of rapid mixing was obtained. An additional reaction time of about five minutes was provided in the tubing connecting the mixer with the filter column, the distribution box, and the water column above the filter bed. The polyelectrolyte used was a cationic quaternary ammonium compound,

Cat-Floc, a product of Calgon Corp., Pittsburg, Pa. The structural formula of the polymer, as reported by the manufacturer is:



Solids reduction was determined using turbidity measurements of water samples withdrawn from various locations along the filter columns. Headloss profiles were determined periodically using a series of piezometers. Each filtration run was terminated when either the headloss reached six feet or the effluent turbidity exceeded 0.2 JTU.

Jar tests were carried out in order to determine the desirable doses of chemicals for flocculation. The tests were performed using a standard multistirrer apparatus and 15 minutes of stirring and 10 minutes of settling were provided. Supernatant samples were withdrawn and analyzed for turbidity.

Four different arrangements of filter media, shown in Figure III-1, were tested. Filter No. 1 had a single media bed whose characteristics were typical of those of full scale filters. The other three filters had various arrangements of dual media beds composed of anthracite and sand. In order to keep the anthracite on top of the sand after backwash the size ratio of the two media must be such that the hindered settling velocity of the sand is not less than that of the coal. According to Conley and Hsiung (Reference III-4) the maximum size ratio is given by:

$$\frac{d_c}{d_s} \leq \left(\frac{S_s - S_m}{S_c - S_m} \right)^{1/1.6} \qquad \text{III-1}$$

Where d_s and d_c are the diameters of the sand and coal respectively. And S_s, S_c, and S_m are the densities of the sand, coal, and the apparent density of the solid-liquid mixture, respectively. If $S_s = 2.60$, $S_c = 1.50$, and $S_m = 1.16$, then Equation III-1 requires that the size of the coal be at

the most 2.47 times that of the sand. In practice the exact size ratio has to be determined experimentally.

Filter No. 2 consisted of 16 inches of anthracite on top of six inches of sand. The anthracite had an effective size, (ES), of 2.5 mm and a uniformity coefficient, (UC), of 1.32. The sand had an ES of 1.10 and a UC of 1.23. Filter No. 3 contained 23-in of coal and 6-in of sand, both of the same type as in Filter No. 2. Filter No. 4 was a dual media filter with both the sand and the coal of finer grade than those used in Filters No. 2 and 3. It contained 16-in of coal ES 0.98 mm, UC 1.35 on top of 6-in of sand ES 0.45 mm, UC 1.55.

Filters No. 2 and No. 3 contained an unusually coarse media. In order to minimize the fine to coarse graduation within each of the media, the fine fractions were screened out. The size distributions of the coal and the sand of these filters are shown in Figure III-2.

The raw water for this series of experiments was prepared as described in Chapter II. The hydraulic loading was 3.0 gpm/sq ft in all but the last five runs in which a loading of 9.0 gpm/sq ft was tested.

EXPERIMENTAL RESULTS

Contact Flocculation with Alum

The desirable alum dose for flocculation-sedimentation was determined by a standard jar test, the results of which are shown in Figure III-3. These data indicate that a dose of about 20 mg/l of alum was appropriate when normal flocculation-sedimentation-filtration was practiced. Contact flocculation tests were carried out with four alum doses ranging from 0 to 20 mg/l. In all cases, the filter runs were terminated because of floc breakthrough. The results are shown in Figure III-4, where the length of run and the terminal headloss are plotted as a function of the alum dose. The longest filter run was obtained with Filter No. 1 using 20 mg/l alum. The best performance of Filters No. 2, 3, and 4 was realized with a 10 mg/l alum dose. As indicated in Figure III-4, only Filter No. 1 developed an appreciable headloss under the various conditions tested. The specific nature of this headloss for the 20 mg/l alum run is shown in Figure III-5 as a function of filter depth for selected time periods. As indicated in the Figure, excessive headloss occurred in the top five inches of the filter bed. This was due to the accumulation of flocs in the top layer causing partial clogging of the filter pores. The same type of flocs could not accumulate in the other three filters which contained coarser media. The early floc

breakthrough and the lack of appreciable headloss for Filters No. 2, 3, and 4 are indicative of a situation where the flocs are not strong enough to form aggregates which can be retained in the large media pores. Therefore, it can be concluded that the flocs formed with alum alone are not suitable for separation by a coarse media filter at the employed flow rate of 3.0 gpm/sq ft.

Contact Flocculation with Polyelectrolytes

Contact flocculation tests employing the cationic polyelectrolyte Cat-Floc were carried out using two different clay concentrations. The polyelectrolyte was used as the sole flocculant in concentrations ranging from 0.5 to 1.0 mg/l for a clay concentration of 6.0 mg/l. In the second series of runs employing 24.0 mg/l clay, the polyelectrolyte dose was increased up to 4.0 mg/l. The primary objective of these experiments was to study the feasibility of using contact flocculation with the very coarse filter media. The results of the tests with 6.0 mg/l clay are illustrated in Figures III-6 through III-8. Under the conditions tested, all the filters demonstrated some capacity for removing suspended solids. However, the required effluent turbidity of 0.2 JTU could not always be achieved. With 0.5 mg/l Cat-Floc, only Filter No. 1 was capable of producing an effluent of acceptable quality. This was due to the accumulation of flocs in the top layers of the sand bed. Also, this accumulation accounts for the rapid buildup of headloss in the filter which lead to the termination of the run after about 10 hrs as the total headloss had reached the limiting value of six ft. As the polymer dose was increased to 0.75 mg/l, acceptable effluents were produced in Filters No. 1 and No. 4, but not in the coarse media Filters No. 2 and No. 3. The length of run in Filter No. 1 was only five hrs compared to the 10 hrs with 0.5 mg/l Cat-Floc. The coarse media Filters No. 2 and No. 3 produced the required low turbidity effluent only with a polymer dose of 1.0 mg/l. These results illustrate the interdependence of floc properties and filter media. For the fine media, Filter No. 1, a Cat-Floc dose of 0.5 mg/l produced a floc of sufficient size and strength such that it was effectively retained by the media. Any increase in polymer dose produced flocs which were too strong for this media and caused increased rates of headloss buildup resulting in shorter filter runs. For the somewhat coarser media, Filter No. 4, a dose of 0.75 mg/l was required to produce the floc of proper quality, and the very coarse media of Filters No. 2 and No. 3 required at least 1.0 mg/l of polymer to produce acceptable results.

Some typical headloss profiles, which developed in the filters with the use of different polymer doses, are shown in Figure III-9. The top graph in Figure III-9 shows the piezometer head profile for Filter No. 1; the filter with the typical single media bed. The major portion of the headloss occurred in the top two to three inches of the bed, as is typical for this type of media where the top layers contain the fine portions of the media. In general, a dual media bed can be expected to exhibit two zones of major headloss: one at the top of the coal or coarse media layer and one at the interface between the coarse and fine media. This expected situation is illustrated in the bottom graph of Figure III which presents the results obtained from Filter No. 4. The sites of floc accumulation in a filter bed are directly correlated to the slope of the piezometric head profile. Thus, those media zones which exhibit a large slope in the piezometric head profile are also the zones where concentrated floc accumulation is occurring. For Filter No. 4, the data of Figure III-9 shows that the floc accumulation at the surface of the anthracite and at the sand-anthracite interface increased with increasing polyelectrolyte additions. Furthermore, the combined data presented in Figures III-6, III-7, III-8 and III-9 indicate that Filter No. 4 did not produce an acceptable effluent until floc started accumulating at the interface. An indication of the reduced floc load placed on the sand layer as a result of the anthracite layer is obtained by noting the lesser slope of the piezometric head profile which prevailed in the sand media of Filter No. 4 and that at the surface of the sand media of Filter No. 1. The results obtained from Filter No. 4 also suggest that the graduation of anthracite in the filter was such that the surface of the anthracite and the sand competed for the same floc at the 0.5 mg/l dosage level of polyelectrolyte. At the higher concentrations of polyelectrolyte, sufficient variation in floc quality was attained such that both the anthracite and the sand contributed to the clarification of the water. However, even for these conditions it appears that the central region of Filter No. 4 accumulated little floc. Also, it is not known if the floc developed at the high polyelectrolyte dosage caused the two zone removal of floc in Filter No. 4, of if the loads placed on the sand resulted from the shearing of the floc collected at the surface of the anthracite layer. The difference between the two mechanisms is important in determining the total load which can be placed on the filter. However, the cause of the difference cannot be evaluated with the data obtained from this study.

The middle graph of Figure III-9 depicts the conditions which prevailed in Filter No. 3 which had large, coarse coal (ES of 2.50 and UC of 1.32) and a large coarse sand (ES of 1.10 and UC of 1.23). The anthracite was sufficiently large so that any fines which did occur at the top of the media did not trap significant quantities of floc. When polyelectrolyte doses of 0.5 and 0.75 mg/l were used, Filter No. 3 (Figures III-6 and III-7) did not provide an acceptable effluent. However, with a polyelectrolyte dose of 1.0 mg/l an acceptable quality of effluent was obtained over an approximately 40 hour period. This duration of run being on the order of 30 hours longer than the runs obtained for Filters No. 1 and 4 under similar conditions (Figure III-8). The piezometric head profile (Figure III-9) shows that little headloss developed in the anthracite layer with most of the headloss being incurred in the six-inch sand layer. However, in this instance the limited headloss in the anthracite layer does not imply little or no floc accumulation in the anthracite, but rather a uniform accumulation of floc through the layer causing a minimum restriction to flow through the layer. The relative importance of the anthracite in accomplishing floc removal without causing rapid headloss buildup is demonstrated in Figure III-8. Comparison of the headloss as a function of time, curves for Filters No. 2 and 3 shows that Filter No. 2, which had seven inches less anthracite than Filter No. 3, experienced a more rapid development of headloss than Filter No. 3. The difference in the headloss between these two filters at any given time is related to the heavier floc load placed on the sand portion of Filter No. 2, as a result of the smaller coal layer which it had. The relative load of floc placed on the anthracite and sand layer in Filters No. 2 and 3 appear to be a function of the quality of floc formed, rather than the consequence of floc removal and floc shearing as was the possible situation for Filter No. 4.

It is quite apparent that when the proper floc properties are obtained, the most desirable filter bed is that which is composed of coarse anthracite and coarse sand. This type of media results in a minimum headloss buildup while still providing adequate turbidity removal.

Floc Breakthrough During the Initial Phase of Filter Run

As can be seen in Figures III-6 through III-8, a certain operating period was required before acceptable effluent could be obtained. This initial breakthrough phase of filter operation has been observed by other

investigators (Reference III-5). Hudson (Reference III-6) suggests that some kind of ripening process is necessary before a filter bed can effectively collect the incoming flocs. He also noted that the ripening period is longer for coarser media. This is in agreement with the results of this study. For the coarse coal filters the initial breakthrough period is quite long. If the duration of this type of breakthrough cannot be reduced, then the feasibility of this type of filter is highly questionable. Wasting the effluent during the ripening period or returning the effluent to the source are possible, but undesirable, solutions. Therefore, an attempt was made to find other techniques for shortening the ripening period. It was assumed that a high dose of alum, applied during the initial phase of the run, would enhance the formation of floc mass coating the media particles; thus reducing the ripening period. Figure III-10 shows the effluent turbidities of Filters No. 2 and 3 as a function of filtration time for three different chemical feeds. When 20 mg/l of alum and no polyelectrolyte was used, floc breakthrough occurred very rapidly. The alum floc was not strong enough to be retained by the coarse media. With the addition of 1.0 mg/l of Cat-Floc an acceptable effluent was obtained after a ripening period of about seven hours. The combination of 20 mg/l alum with 1.0 mg/l Cat-Floc produced an acceptable effluent immediately after the start of the run. The alum floc is strengthened by the polymer, thus eliminating early alum floc breakthrough. The large mass of alum floc provides the necessary coating to the media which makes it very effective in collecting the incoming flocs. The alum addition was stopped after five hours of operation to avoid excessive clogging which would result in shorter filter runs. The addition of alum in the first five hours caused only a minor reduction in the total duration of the filter runs.

The termination of alum addition after five hours resulted in an immediate and significant increase in effluent turbidity. The turbidity level in the effluent quickly reached a maximum then slowly declined as the filter run proceeded. The reason for this shift in turbidity was not pursued, although it is reasonable to assume that it was the result of shifts in chemical equilibria which had to be developed as a result of the elimination of alum from the influent water. However, the practicality of using alum to reduce the initial breakthrough phase is demonstrated, and the shift in effluent quality resulting from the termination of alum addition does not nullify the value of the approach.

Contact Flocculation of Water of Higher Turbidity

A series of filtration runs on water containing 24 mg/l clay was conducted in an effort to test the effect of higher clay concentrations on the length of filter run and on the required chemical dose. The results are summarized in Figure III-11, showing the length of filter run and the cause of termination. The interrelation between floc properties, filter media, and filter performance are implicit in these results. For the fine media of the type in Filter No. 1, a dose of 1.0 mg/l Cat-Floc is sufficient to produce floc which is effectively retained by the upper portion of the bed and terminate the filter run by reaching the limiting six ft of headloss. Any increase in polymer dose results in the formation of a still stronger or larger floc, thus shortening the filter run. Slightly coarser media, such as used in Filter No. 4, exhibits a similar behavior to that of Filter No. 1 with somewhat longer runs due to the reduced headloss associated with flow through the coarser media at any given level of floc retention. With a very coarse media of the type and form used in Filters No. 2 and No. 3, a low polymer dose results in floc breakthrough because the flocs are not sufficiently strong or large to be retained by the coarse media. As long as the filter run is controlled by floc breakthrough the length of run will increase with polymer dose. The headloss also increased with polymer dose, and as the dose reached 4.0 mg/l the headloss became limiting. Polymer doses higher than 4.0 mg/l were not used, but it is thought that shorter filter runs would have resulted with any polymer dose above 4.0 mg/l. Therefore, it seems that for the coarse media there is a certain polymer dose which yields the longest filter run, and that with this dose the limiting headloss and the floc breakthrough might occur at the same time. This polymer dose is not necessarily the economically optimum dose, as the economy of filter operation depends on many other considerations which may indicate other more efficient operating conditions.

The headloss profiles at the termination of the filter runs are shown in Figure III-12 for the different filters and different polymer doses. The figure illustrates the effect of floc penetration on the distribution of headloss along the filter bed and its impact on filtration efficiency. Floc penetration in the fine media beds of Filters No. 1 and No. 4 floc was limited to the upper few inches of the bed resulting in an excessive local headloss. The exception was the 1.0 mg/l of Cat-Floc run in Filter No. 4 where floc removal occurred throughout the anthracite layer. The problem of surface layer removal

is intensified as the polymer dose increases. The nature of the removal process which existed in Filter No. 3 was quite different from that in Filters No. 1 and 4. For the coarse media, as used in Filter No. 3, rapid floc breakthrough and low overall headloss are associated with low polymer doses. Polymer doses of 2, 3, and 4 mg/l resulted in floc retention throughout the coal layer. The characteristics of the floc were such that breakthrough was delayed, giving rise to longer filter runs. The longest filter run was obtained with 4.0 mg/l Cat-Floc as more of the filter bed was utilized. For this particular situation the headloss was the limiting factor.

Contact Flocculation at High Flow Rates

A series of experimental filter runs was conducted at a flow rate of 9.0 gpm/sq ft. Because of limited hydraulic capacity, only one filter could be operated at this high hydraulic loading rate. The tests were carried out using the deep and coarse dual media in Filter No. 3. In an attempt to prevent initial breakthrough, 20 mg/l of alum were added during the first two hours of each run. However, as the results indicate (Figures III-13 through III-17) the alum addition retarded, but did not eliminate, early breakthrough. Based on the data of Figures III-13, III-15, III-16, and III-17, the termination of alum addition appears to have resulted in an immediate and significant decrease in effluent turbidity. This relationship between cessation of alum addition and effluent turbidity is essentially the opposite of that observed at lower flow rates, Figure III-10. For these high rate runs, an acceptable quality effluent was obtained for a period of six to eight hours after the early breakthrough.

Comparison of the results presented in Figure III-13 with those presented in Figure III-14 indicates that the use of alum intensified the severity of the breakthrough. An additional effect of alum on the filtration process is illustrated in Figure III-18, these data having been obtained from the same runs as the data shown in Figures III-13 and III-14. It is apparent from the results presented in Figure III-18 that the alum caused a more effective use to be made of the coal layer of the filter. When alum was added during the first two hours of the run, the effectiveness of utilization of the coal layer was increased by approximately 60 percent. The use of alum appears to produce a coal-alum complex which is more effective in removing the floc generated by the addition of polymer than is the coal alone. Further, the data suggests that the coal has a limited capacity to contain the alum, and once that capacity is satisfied the alum floc is sheared and readily penetrates

both the coal and sand media. Additional experiments, which were beyond the limits of this study, should establish an optimum duration for the addition of alum for high rate filtration.

DISCUSSION OF RESULTS

The experimental results obtained in the contact flocculation phase of the study are summarized in Table III-1. The results are expressed in terms of the length of run and the cause of termination, that is either floc breakthrough or excessive headloss. The experiments demonstrated that with the proper combination of filter media, chemical addition, and operating procedure, a contact flocculation system can effectively remove suspended solids from turbid water, and produce a high quality effluent in what seems to be a very economic process. The limited number of experiments performed could not provide the data necessary to examine all aspects of this complicated process. However, a number of interesting interrelationships and trends can be deduced from the data obtained.

Perhaps the most important conclusion that can be drawn from the results is the interdependence of all operating and performance parameters. Thus, the concepts of optimum media, optimum flow rate, or optimum chemical dose are at variance with the manner in which a filter performs and responds. There is only an optimum combination of media, flow rate, and chemical dose which will offer, for a given raw water, the most economical clarification system. The selection of such an optimum combination is not an easy task and is to some extent dependent on the specifics of each situation. However, some general conclusions, derived from the experimental results, can serve as guidelines.

Filter Media

The major problem with most media is that they bring about the accumulation of floc in the top layers of the filter bed. The ideal media is one which would result in a uniform floc and headloss distribution over the depth of the bed under a wide variety of conditions and still produce acceptable effluent.

A single media, fine-to-coarse, filter bed is at variance with the preceding concept. The upper layers of fine sand clog rapidly causing short filter runs. Regular dual media filters, consisting of a well graded coal layer on top of a sand layer, provide a partial solution to the problem. However, a fine-to-coarse graduation of the coal can be expected to result in the formation of high headloss regions in the upper portion of the bed.

TABLE III-1

SUMMARY OF CONTACT FLOCCULATION RESULTS

Run No.	Hydraulic Loading (gpm/sq ft)	Raw Water Suspended Solids (mg/l)	Alum Dose (mg/l)	Cat-Floc Dose (mg/l)	Length of Filter Run, hrs			
					Filter No.1	Filter No.2	Filter No.3	Filter No.4
1	3.0	6.0	5	0	2.0 $_{B}$	0 $_{B}$	0 $_{B}$	0 $_{B}$
2	3.0	6.0	10	0	6.0 $_{B}$	3.5 $_{B}$	4.5 $_{B}$	6.5 $_{B}$
3	3.0	6.0	20	0	7.0 $_{H}$	2.5 $_{B}$	1.5 $_{B}$	3.5 $_{B}$
4	3.0	6.0	0	0.5	10.0 $_{H}$	0 $_{B}$	0 $_{B}$	0 $_{B}$
5	3.0	6.0	0	0.75	5.5 $_{H}$	0 $_{B}$	0 $_{B}$	14.0 $_{H}$
6	3.0	6.0	0	1.0	4.5 $_{H}$	33.0 $_{H}$	43.0 $_{H}$	9.5 $_{H}$
7	3.0	24.0	0	1.0	2.5 $_{H}$	0.5 $_{B}$	0.5 $_{B}$	10.0 $_{H}$
8	3.0	24.0	0	2.0	2.0 $_{H}$	6.0 $_{B}$	11.5 $_{B}$	4.0 $_{H}$
9	3.0	24.0	0	3.0	1.5 $_{H}$	9.5 $_{B}$	14.5 $_{B}$	2.5 $_{H}$
10	3.0	24.0	0	4.0	1.3 $_{H}$	11.0 $_{H}$	14.5 $_{H}$	2.3 $_{H}$
11	9.0	6.0	0	1.0	-	-	10.0 $_{H}$	-
12	9.0	6.0	0	2.0	-	-	8.0 $_{H}$	-
13	9.0	6.0	0	3.0	-	-	8.5 $_{H}$	-
14	9.0	6.0	0	4.0	-	-	9.0 $_{H}$	-

NOTES:

B = run terminated by floc breakthrough.

H = run terminated by limiting headloss.

The problem, based on the results of this study, is almost completely resolved by using a coarse and uniform coal layer as the upper part of a dual media filter. The results, summarized in Table III-1, substantiate these conclusions. The relative advantages of the three filter media combinations can be determined by comparing the length of filter runs under conditions of constant flow rate and raw water quality. The data for such a comparison are presented in Table III-2. The data presented in Table III-2 are for the three filters which all had 22-inches of media and differed only in the composition of the filtering media.

TABLE III-2

EFFECT OF FILTER MEDIA ON LENGTH OF RUN

Filter No.	Media	Longest Filter Run, Hrs	
		With 6.0 mg/l Suspended Solids	With 24.0 mg/l Suspended Solids
1	Single Media, Sand	10	2.5
4	Dual Media, Fine Coal and Fine Sand	14	10
2	Dual Media, Coarse Coal and Coarse Sand	33	11

Conditioning and Chemical Dose

The experimental results indicate that polyelectrolyte is far superior to alum as a filter aid in contact flocculation, irrespective of the type of media used. In the typical fine-to-coarse sand filter, the volume of floc produced with alum is larger than that produced by polyelectrolytes and causes rapid clogging of the fine top sand layers. The alum floc is not strong enough to be retained in the large pores of the coarser dual media filters; thus, early breakthrough results in short filter runs. The polyelectrolyte flocs are tenacious enough to be retained in these latter type filters.

In addition to the above technological advantages, the use of Cat-Floc may in many cases be less expensive than alum. Based on an estimated alum cost of 4¢/lb compared to 40¢/lb of Cat-Floc, 1.0 mg/l of the polymer is the cost equivalent to 10.0 mg/l alum. However, in terms of flocculating power, 1.0 mg/l Cat-Floc may be equivalent to more than 10 mg/l alum. In the present study, the use of 1.0 mg/l Cat-Floc resulted in filter performance far exceeding those obtained with 20 mg/l alum. Also, additional savings can be

expected in smaller chemical storage and application facilities. It should be noted that at the present time Cat-Floc is the only cationic polyelectrolyte approved by the United States Public Health Service for use in domestic water supplies.

The optimum chemical dose depends on the filter media, the suspended solids contents of the raw water, and probably on the flow rate. In the sand filter and the fine dual media filter, where fast clogging is the most probable cause of filter run termination, a small chemical dose is required. The polymer dose seems to be in direct proportion to the concentration of suspended matter. In recognition of the concept that polymers serve as bridging material for binding the suspended particles, this result would be expected. The data supporting the preceding conclusions are summarized in Table III-3.

TABLE III-3

EFFECT OF CHEMICALS ON LENGTH OF RUN

Filter No.	Media	6.0 mg/l Suspended Solids				24.0 mg/l Suspended Solids	
		With Alum		With Cat-Floc		With Cat-Floc	
		Dose mg/l	Longest Run, hrs	Dose mg/l	Longest Run, hrs	Dose mg/l	Longest Run, hrs
1	Sand	20	7.0	0.5	10.0	1.0	2.5
2	Coarse Dual Media	10	3.5	1.0	33.0	4.0	11.0
3	Coarse Dual Media (Deep Coal Layer)	10	4.5	1.0	43.0	4.0	14.5
4	Fine Dual Media	10	6.5	0.75	14.0	1.0	10.0

Raw Water Suspended Solids Contents

The concentration of suspended solids in the raw water affects the system's operation in several ways. The polyelectrolyte dose for the coarse filter media appears to be proportional to the raw water suspended solids concentration. Thus, as can be seen from Table III-3, Filters No. 2 and No. 3 required 1.0 mg/l Cat-Floc for 6.0 mg/l suspended solids and 4.0 mg/l Cat-Floc for 24.0 mg/l suspended solids. For the sand filter and the fine dual media filter, the polyelectrolyte dose required for 24.0 mg/l of suspended solids was only slightly higher than that needed for 6.0 mg/l of suspended solids.

In all cases, shorter filter runs were obtained with the higher concentration of suspended solids. This expected observation has a theoretical explanation. According to Ives (Reference III-7), the hydraulic gradient along a filter column is related to the volume of the deposited flocs by

$$\frac{dh}{dl} = \left(\frac{dh}{dl}\right)_o + K\sigma \qquad \text{(III-2)}$$

where h is the piezometric head, k is a headloss constant, σ is the volume of floc retained per unit of filter volume, and $\left(\frac{dh}{dl}\right)_o$ is the headloss per unit of filter length at the beginning of a run. If K is assumed to be independent of depth, Equation (III-2) can be integrated to give the total headloss across the filter; thus,

$$H = H_o + K\int_o^L \sigma dl \qquad \text{(III-3)}$$

where H_o and H are the total headlosses before and after floc accumulation, respectively, and L is the filter depth. The total volume of floc retained by the filter is given by

$$V_F = A\int_o^L \sigma dl \qquad \text{(III-4)}$$

where A is the cross sectional area of the filter bed. Substituting Equation (III-4) into Equation (III-3) yields

$$H = H_o + \frac{K}{A} V_F \qquad \text{(III-5)}$$

Therefore, for a specific media and flow rate, the total headloss depends only on the volume of floc retained by the filter. This leads to the conclusion that, for a given filter at constant flow rate, the time to reach a fixed headloss depends only on the volume of floc formed from the suspended solids in the raw water and the added chemicals. It should be noted that Equation (III-5) implies that the distribution of flocs along the filter depth has no effect on the total headloss and this is in contradiction to many experimental observations showing that deeper penetration of floc results in less overall headloss.

For the coarse media filters, where floc penetration is substantial, the length of filter run is expected to be inversely proportional to the concentration of floc in the raw water. With 24.0 mg/l suspended solids, the floc volume obtained with polyelectrolyte will be approximately four times as great as with 6.0 mg/l. Hence, filter runs are expected to be 25 percent as long. The experimental results show (Table III-3) that the length of run for the low suspended solids concentration was three times longer than that for the 24.0 mg/l suspended solids concentration. The agreement between concept and observation is relatively good.

Rate of Filtration

The effect of flow rate on the length of a filter run can be deduced from an analysis of Equation (III-5). V_F is proportional to the product of raw water suspended solids, C_o, the flow rate, q, and the length of run, T_H, which is the time required to reach the limiting headloss. Thus, using a proportionality constant K_1, the volume of floc is

$$V_F = K_1 \, q \, C_o \, T_H \qquad \text{(III-6)}$$

Substituting this relationship into Equation (III-5) and rearranging yields

$$T_H = \frac{(H-H_o) \, A}{K \, K_1 \, C_o \, q} \qquad \text{(III-7)}$$

In this equation both H_o and K are flow rate dependent. For Filter No. 3 the following values of H_o were measured:

at; $q = 3.0$ gpm/sq ft; $H_o = 0.9$-in

$q = 9.0$ gpm/sq ft; $H_o = 5.5$-in

The variation of K with q is not known. However, it can be estimated by comparing the results at the two flow rates tested. As A, K_1, and C_o are assumed to be independent of q,

$$\frac{T_H q}{(H-H_o)} = \frac{K_2}{K} \qquad \text{(III-8)}$$

where K_2 is a constant independent of flow rate. Substituting the following values into Equation (III-8) yields:

	q gpm/sq ft	H_o in.	H in.	T_H hrs.
Low Flow Rate	3.0	0.9	70	43.0
High Flow Rate	9.0	5.5	70	10.0

$$\frac{K \text{ at } 9.0 \text{ gpm/sq ft}}{K \text{ at } 3.0 \text{ gpm/sq ft}} = 1.33$$

Thus the product $q \, T_H$ decreases with increasing q. This means that the volume of water which can be filtered during a single filter run decreases as the flow rate increases. This is to be expected as the Reynolds numbers at 3.0 and 9.0 gpm/sq ft are about 12 and 36, respectively, and H_o is not linearly related to q.

Depth of Filter Bed

Some indication of the effect of filter depth on filter performance can be obtained by comparing the operation of Filters No. 2 and 3. These filters

were of the same media, differing only in depth. Filter No. 2 had 16-in of coal compared to the 23-in in Filter No. 3. Both had an identical 6-in sand layer at the bottom. The characterisitics of the filter and performance data are compared below:

	Filter No. 2	Filter No. 3	Ratio
Depth of Coal Layer	16-in	23-in	1.44
Total Depth	22-in	29-in	1.32
Length of Run with 6.0 mg/l Suspended Solids	33.0 hrs	43.0 hrs	1.30
Length of Run with 24.0 mg/l Suspended Solids	11.0 hrs	14.5 hrs	1.32

Although the length of filter run in these experiments varied linearly with the total filter depth, it is not suggested as a conclusion that this can be generalized for other filters or other depths. It merely reflects the fact that deeper coal layers have larger capacities for retaining suspended matter without experiencing excessive headloss.

For the dual media type of filter, Equation (III-5) can be written as,

$$H-H_o = \left[\frac{K}{A} V_F\right]_c + \left[\frac{K}{A} V_F\right]_s \qquad \text{(III-9)}$$

where the subscripts c and s refer to coal and sand respectively. If $K_c \ll K_s$, it is advantageous to have more of the flocs retained in the coal portion of the filter. This can be achieved only by increasing the depth of the coal layer. On the basis of this simplified analysis, it is expected that the length of filter run is proportional to some "equivalent" depth which depends on the physical depth of each layer and the K coefficient of the layer. In the particular example listed above, the equivalent depth happens to be equal to the physical depth. Until more evidence is obtained this should not be regarded as the general case.

FILTER MEDIA FOR CONTACT FLOCCULATION

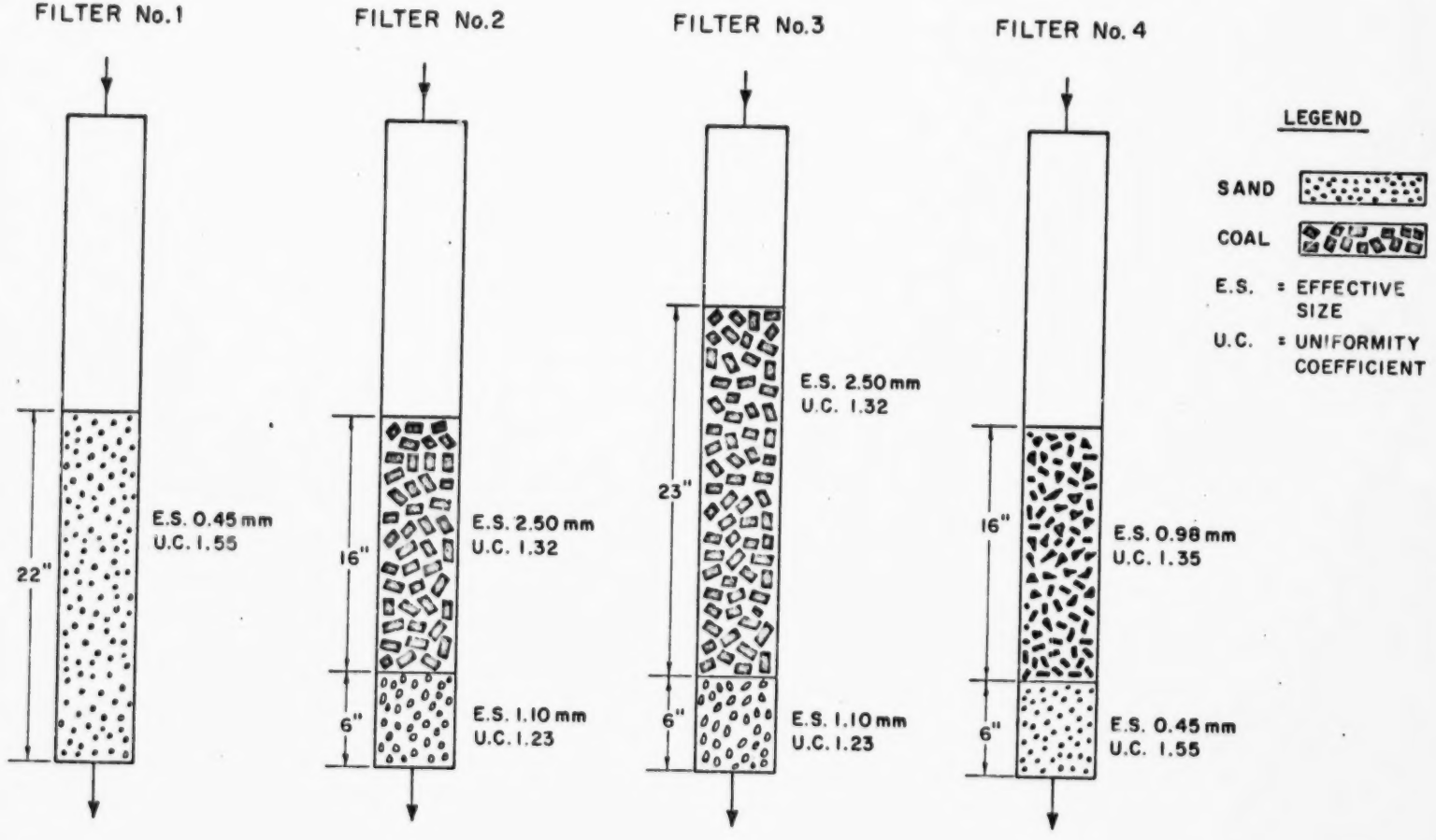


FIGURE III-1

III-19

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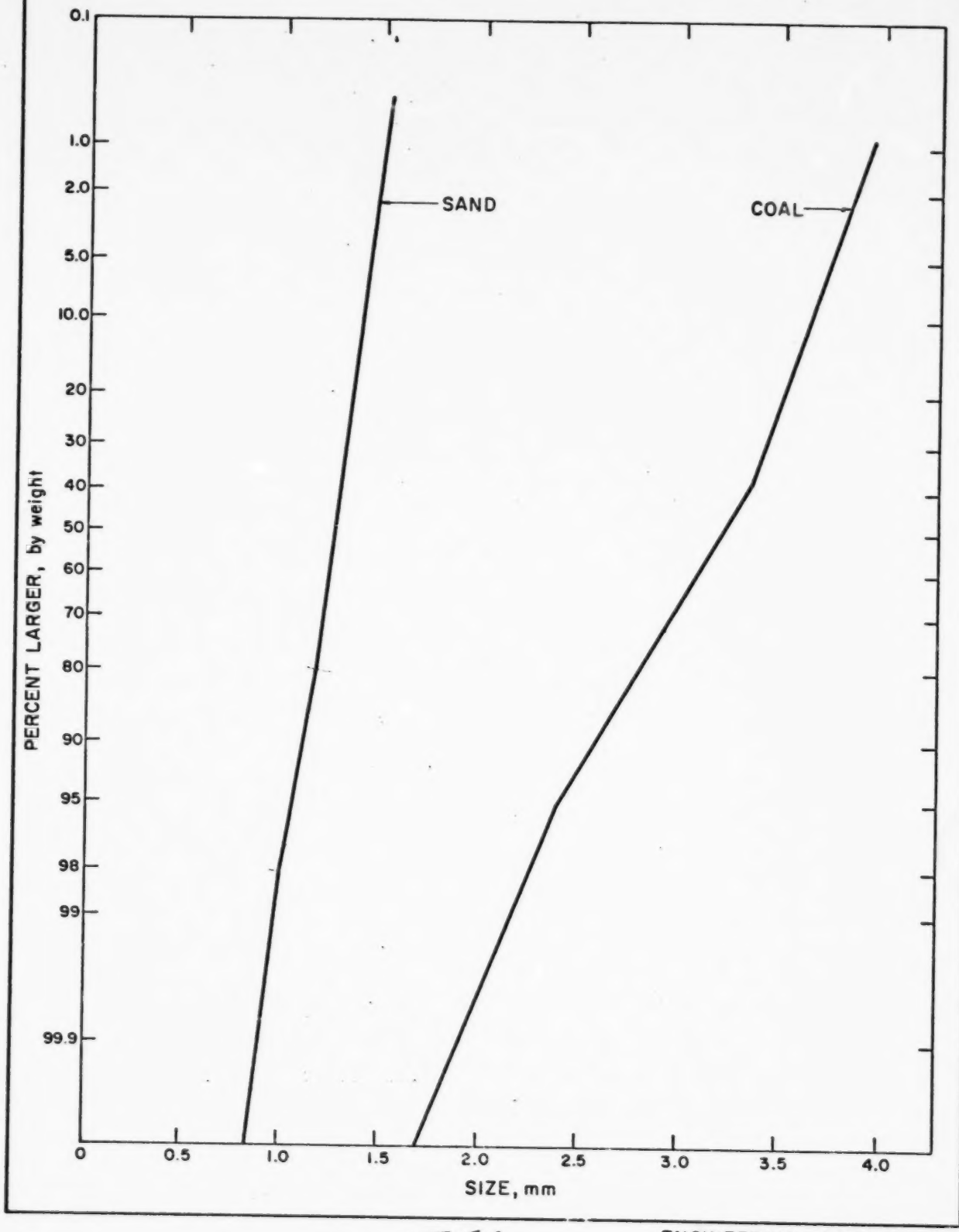
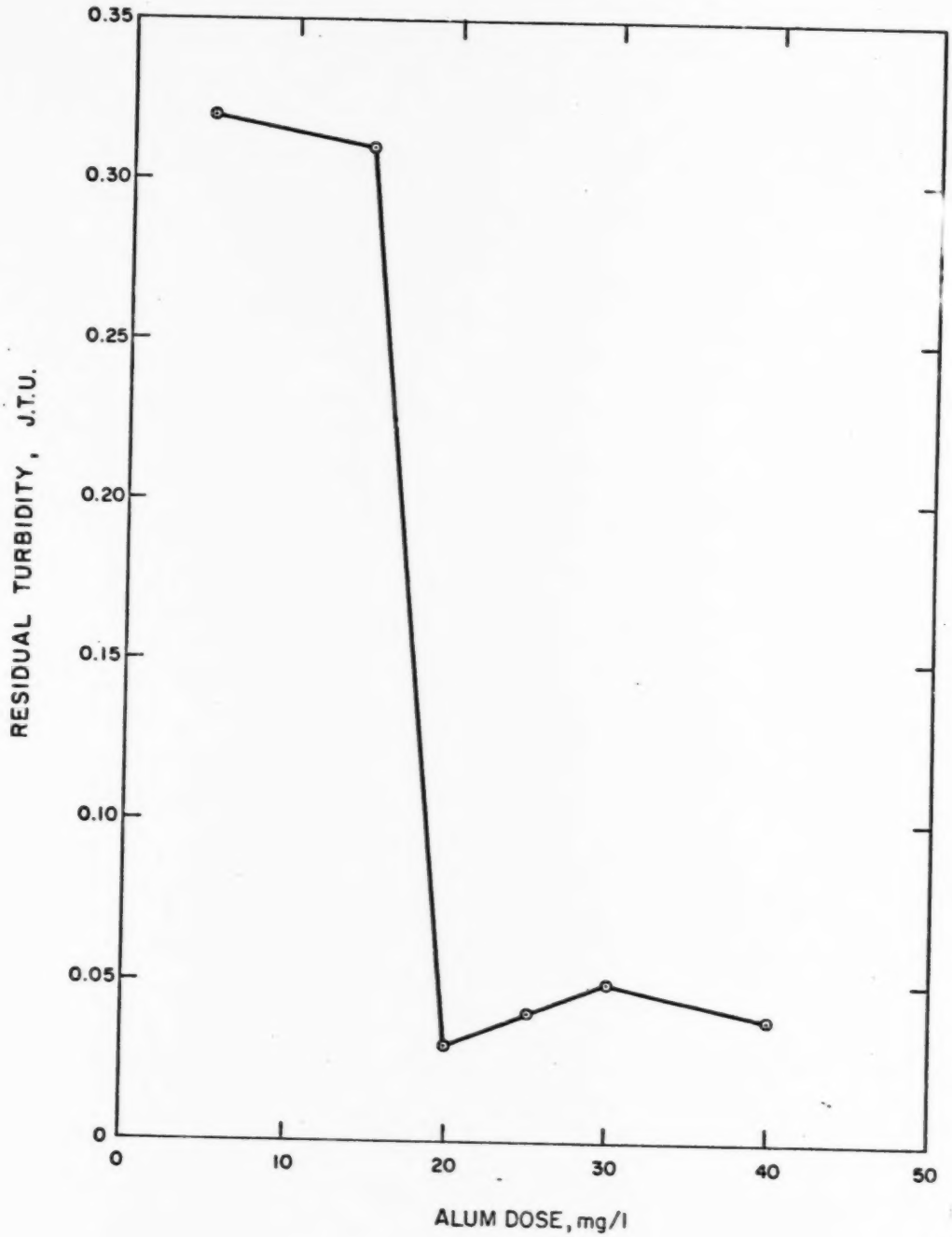
FIGURE III-2

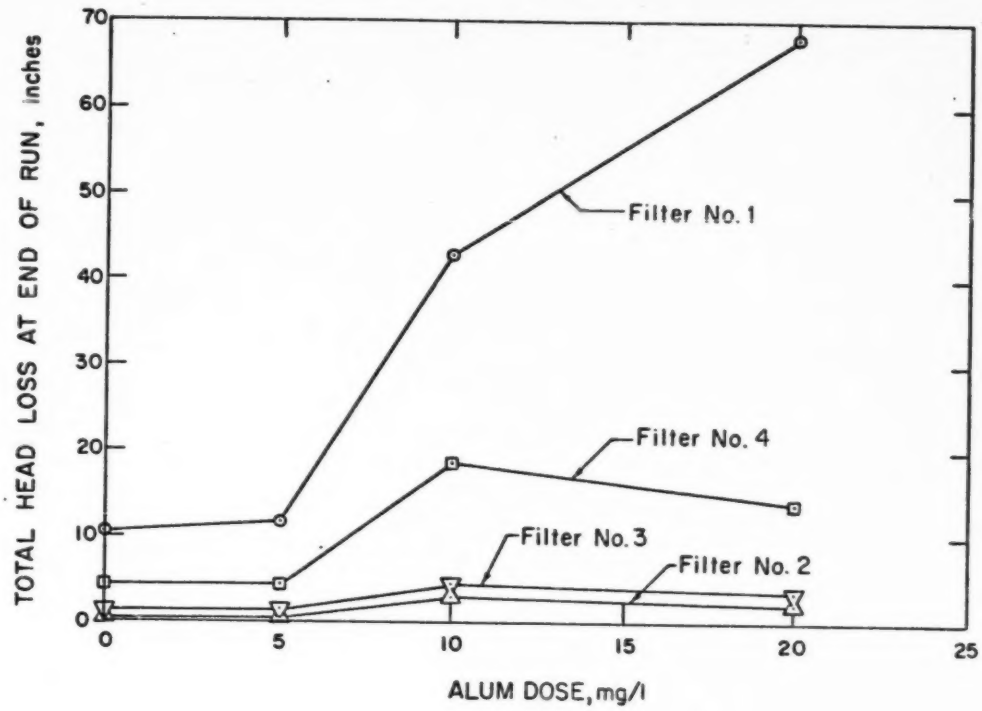
SIZE DISTRIBUTION OF MEDIA FOR FILTERS No.2 AND No.3

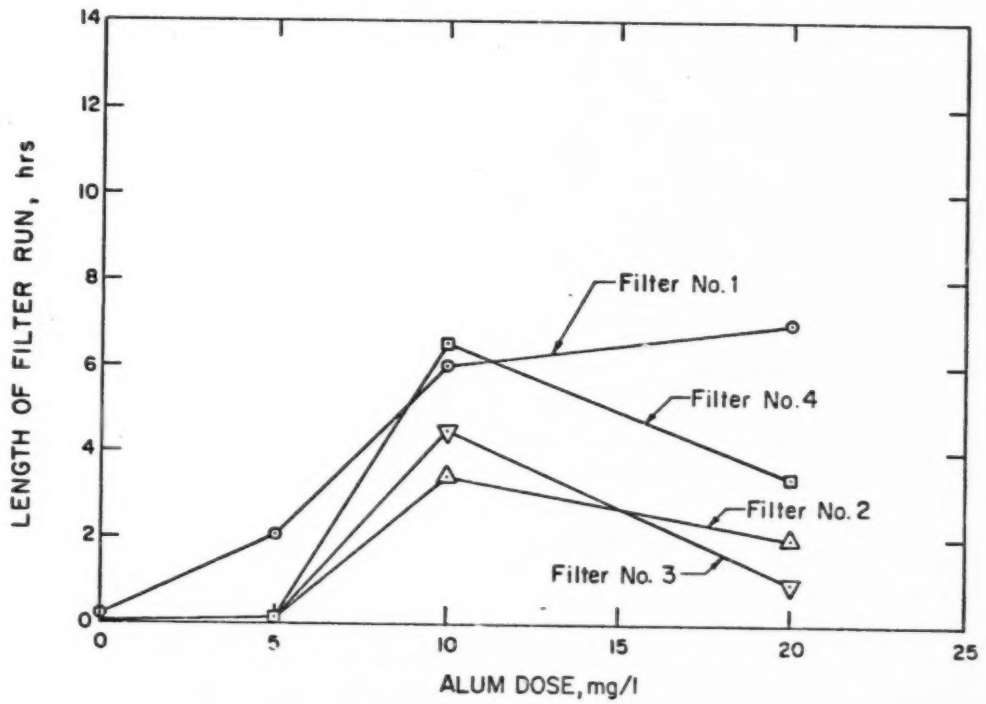


RESULTS OF JAR TEST WITH ALUM

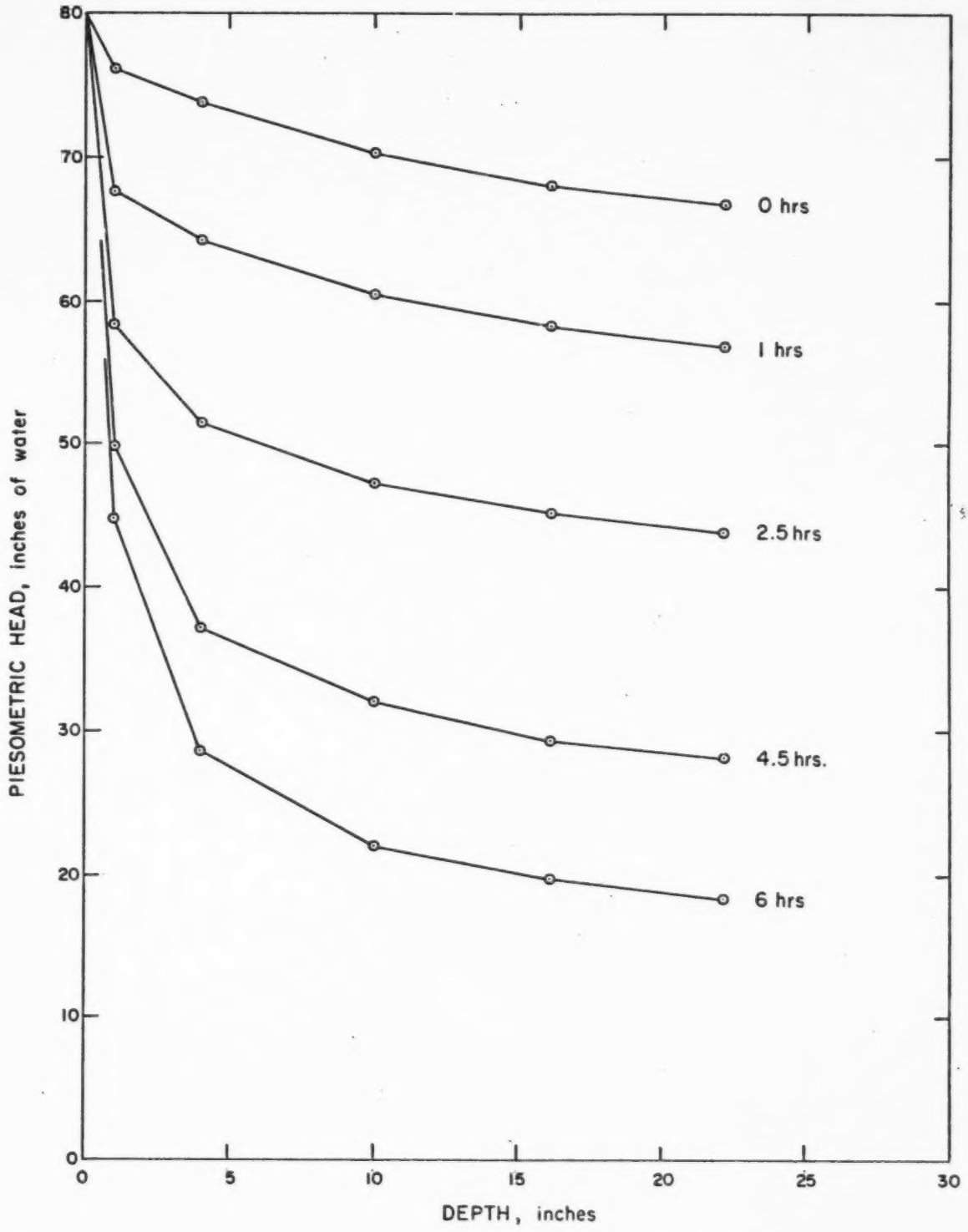


CONTACT FLOCCULATION USING ALUM
6.0 mg/l CLAY, 3.0 gpm/sq. ft.

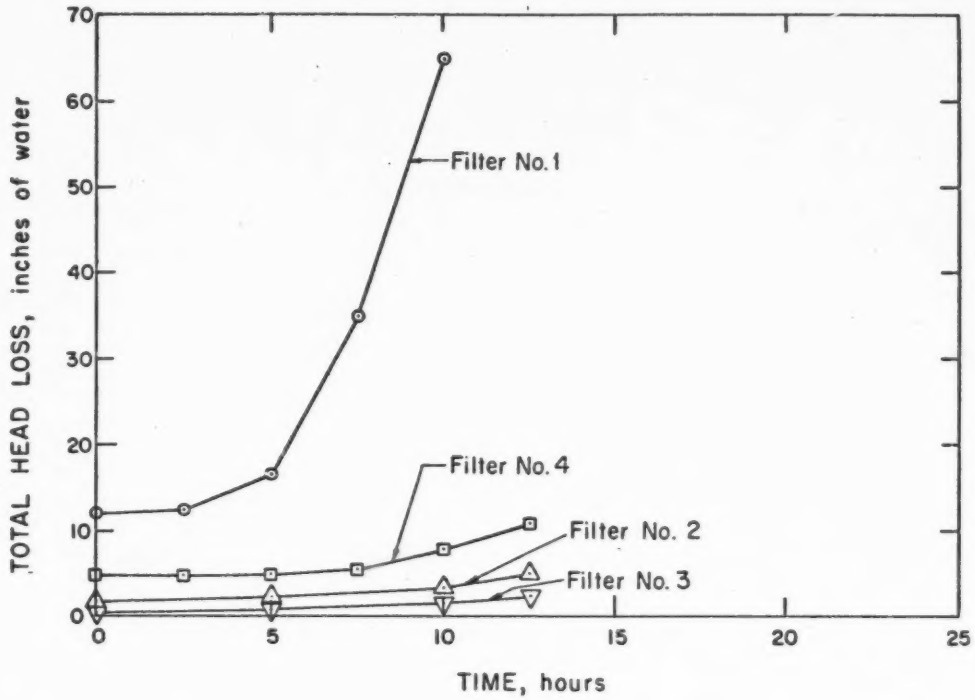


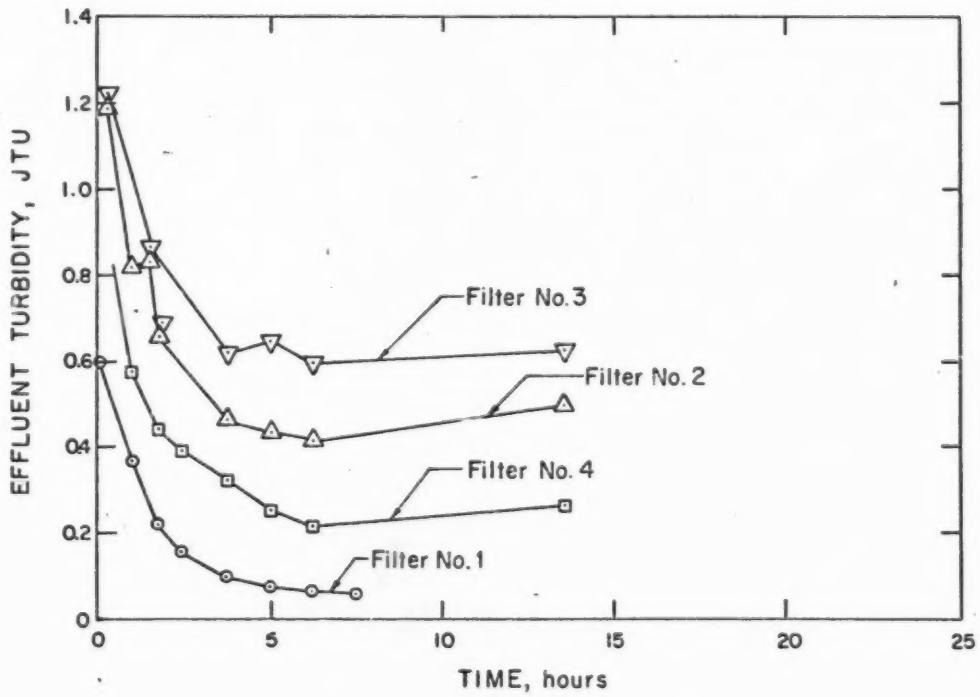


HEAD LOSS PROFILES IN FILTER No. 1
6.0 mg/l CLAY, 20 mg/l ALUM, 3.0 gpm/sq.ft.

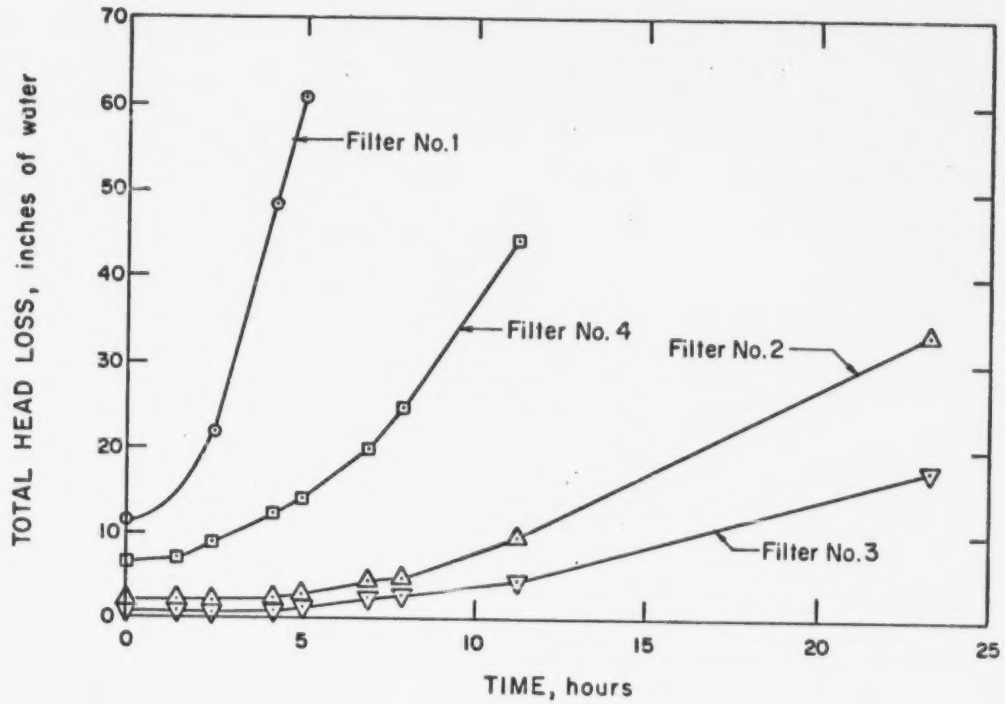


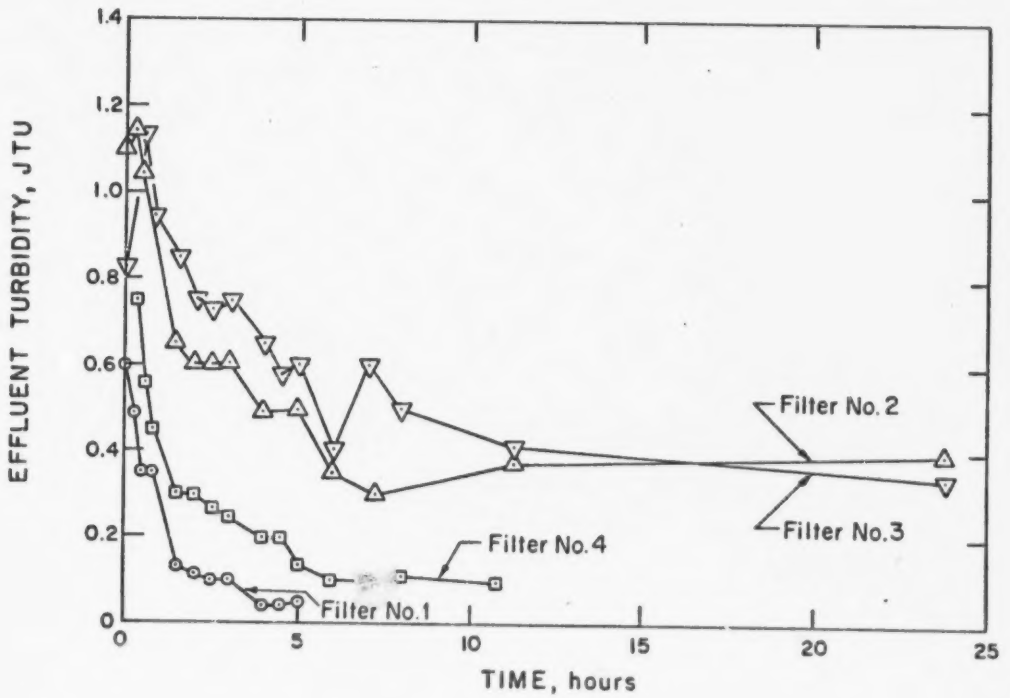
CONTACT FLOCCULATION WITH 0.5 mg/l CAT FLOC, 6.0 mg/l CLAY, 3.0 gpm/sq.ft.



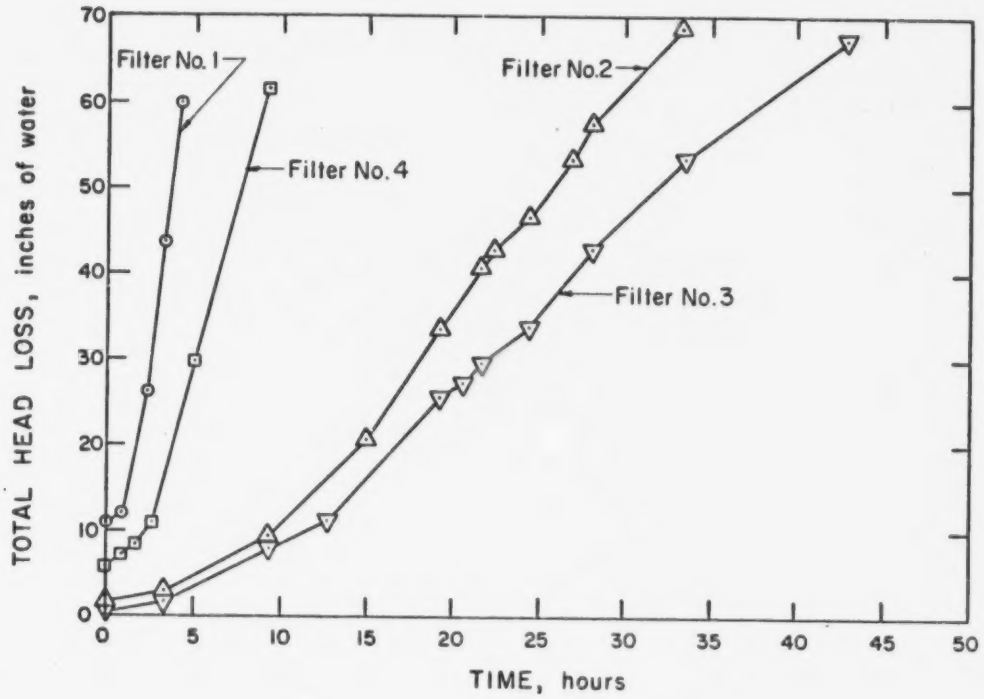


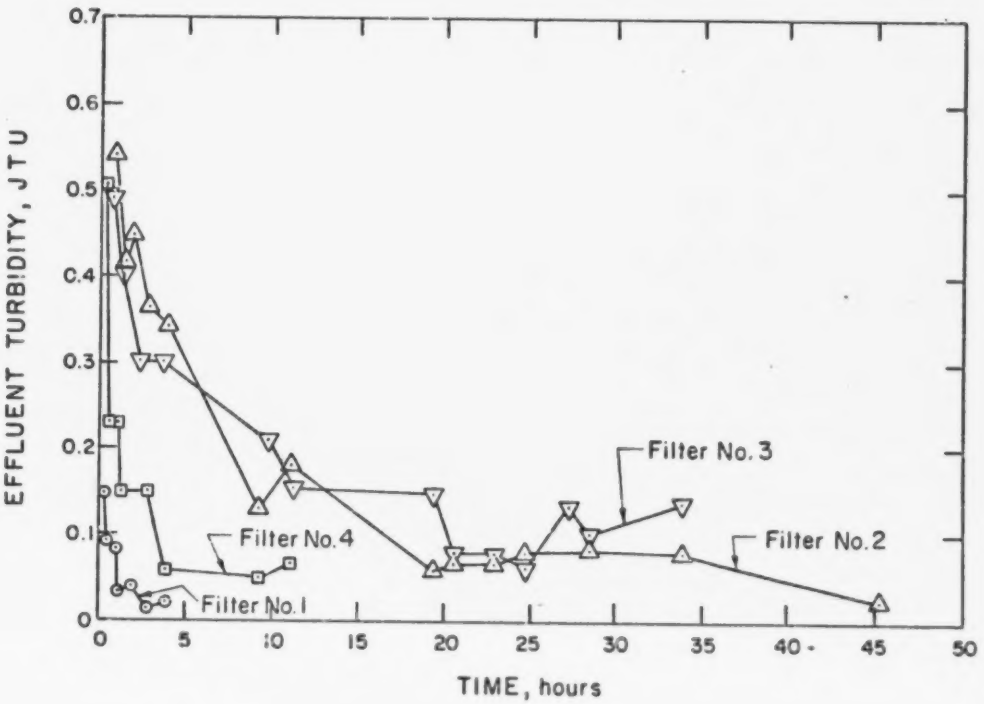
CONTACT FLOCCULATION WITH 0.75 mg/l CAT FLOC, 6.0 mg/l CLAY, 3.0 gpm/sq.ft.



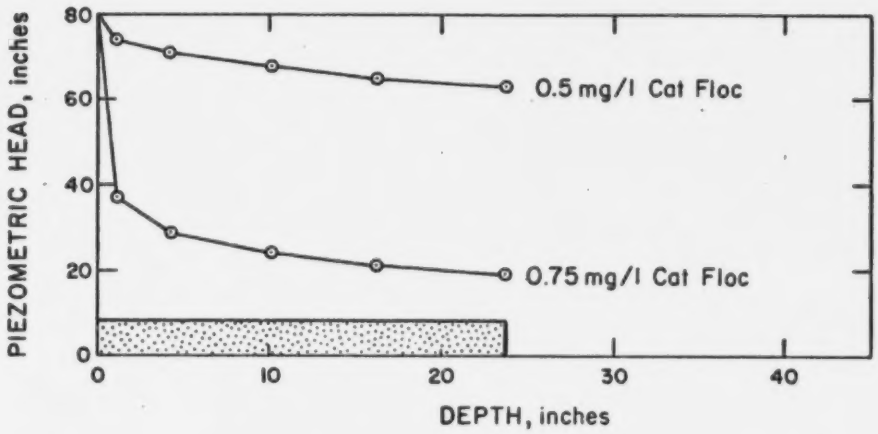


CONTACT FLOCCULATION WITH 1.0 mg/l CAT FLOC, 6.0 mg/l CLAY, 3.0 gpm/sq.ft.

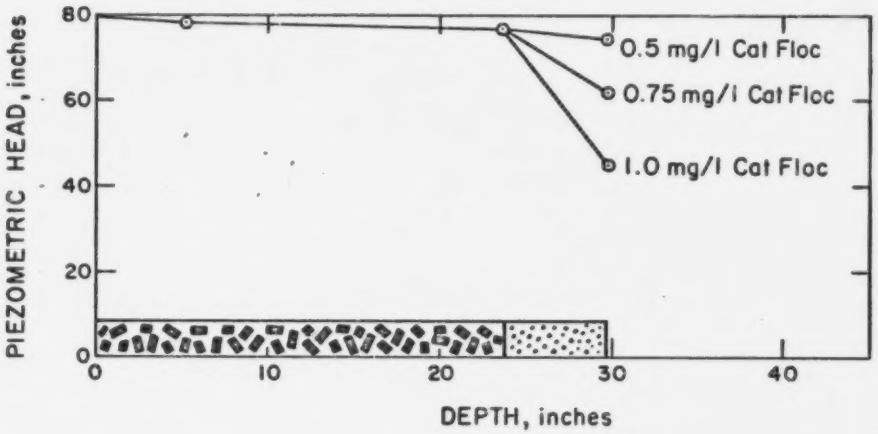




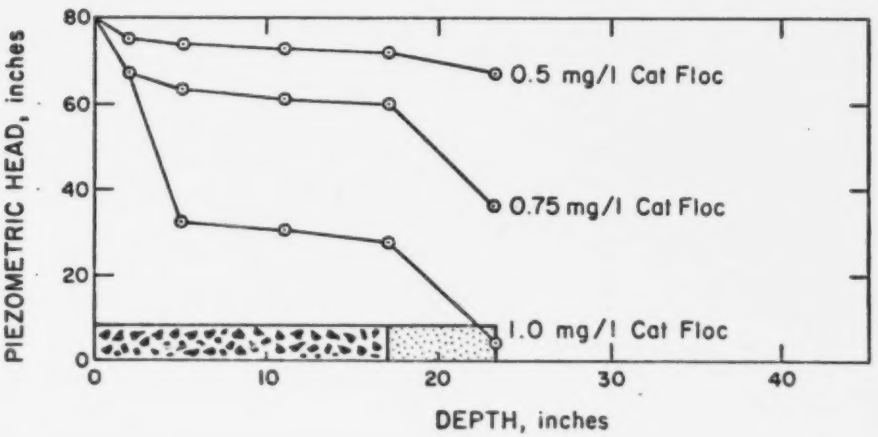
HEAD LOSS PROFILES, CONTACT FLOCCULATION WITH CAT FLOC 6.0 mg/l CLAY, 3.0 gpm/sq.ft.



Filter No. 1 at 5 hours after start of run.



Filter No. 3 at 23 hours after start of run.



Filter No. 4 at 12 hours after start of run.

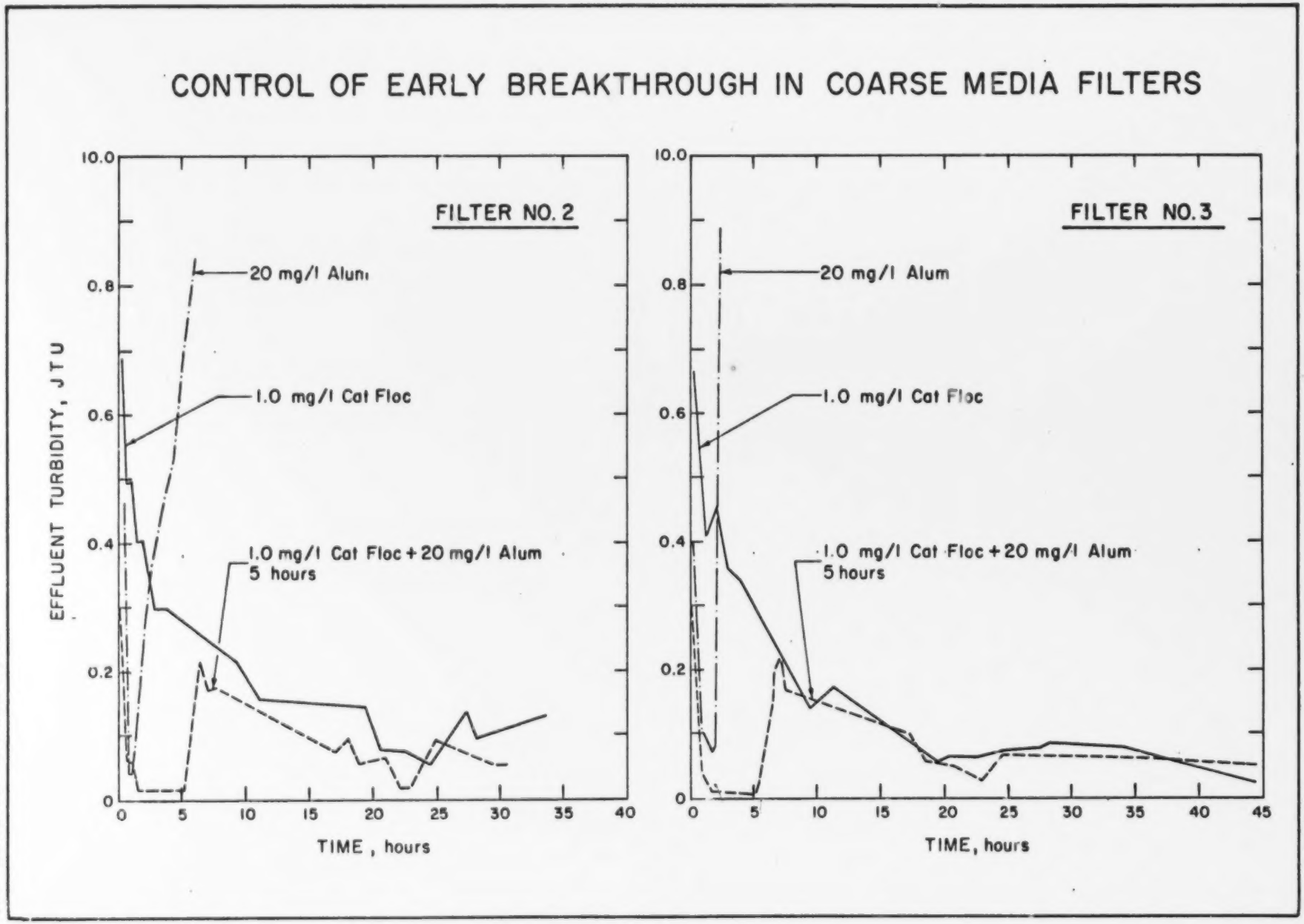
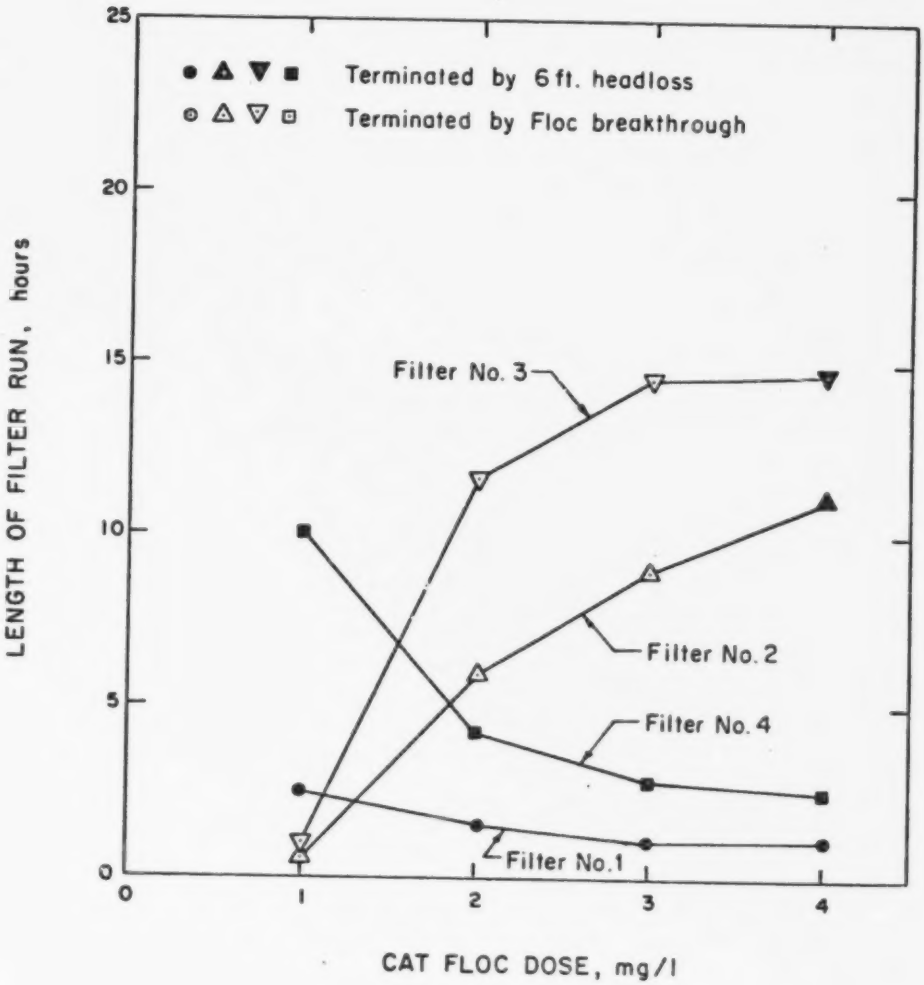


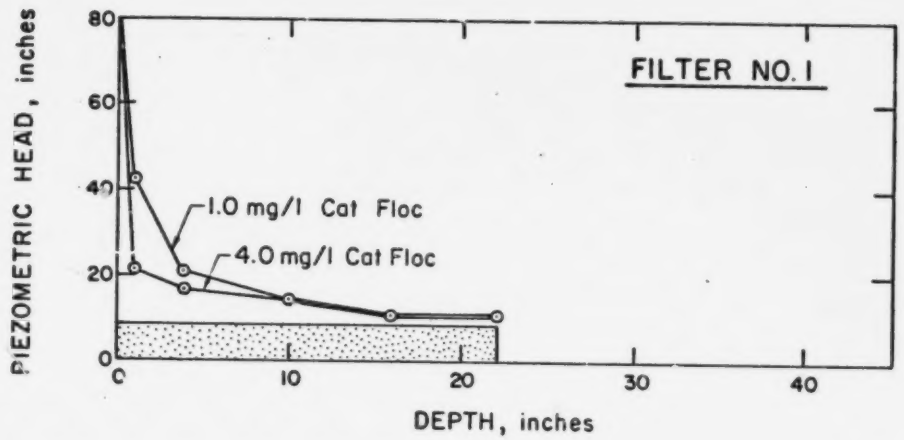
FIGURE III-10

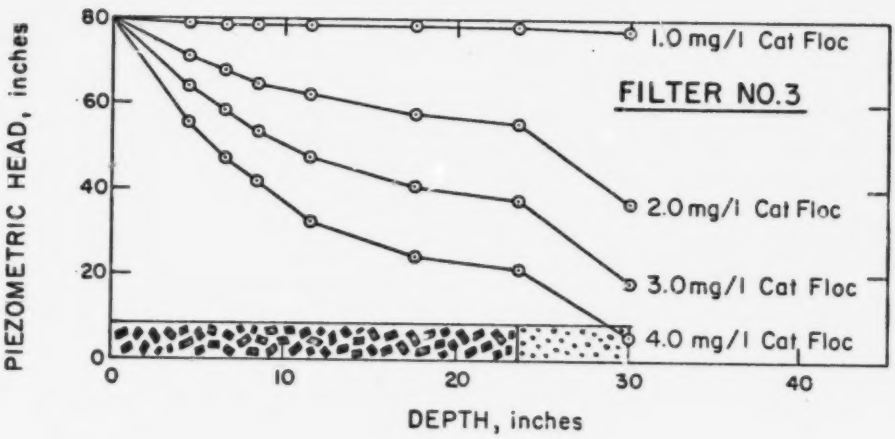
ENGINEERING-SCIENCE, INC.

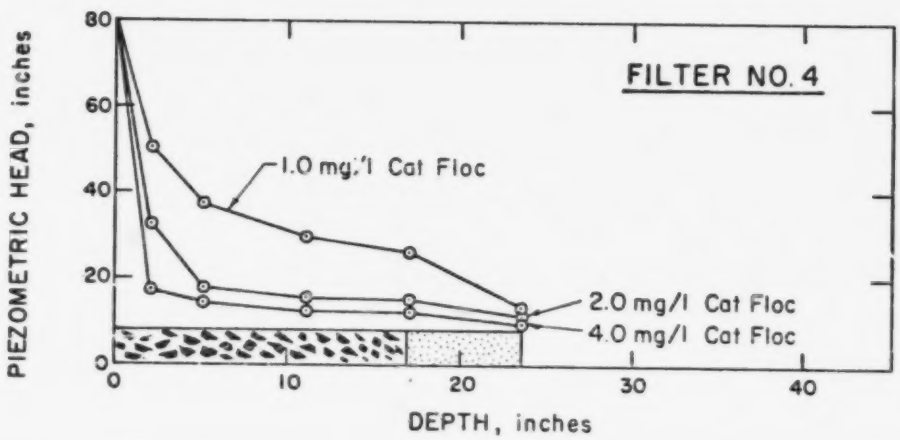
EFFECT OF POLYMER DOSE ON LENGTH OF FILTER RUN 24.0 mg/l CLAY, 3.0gpm/sq.ft.



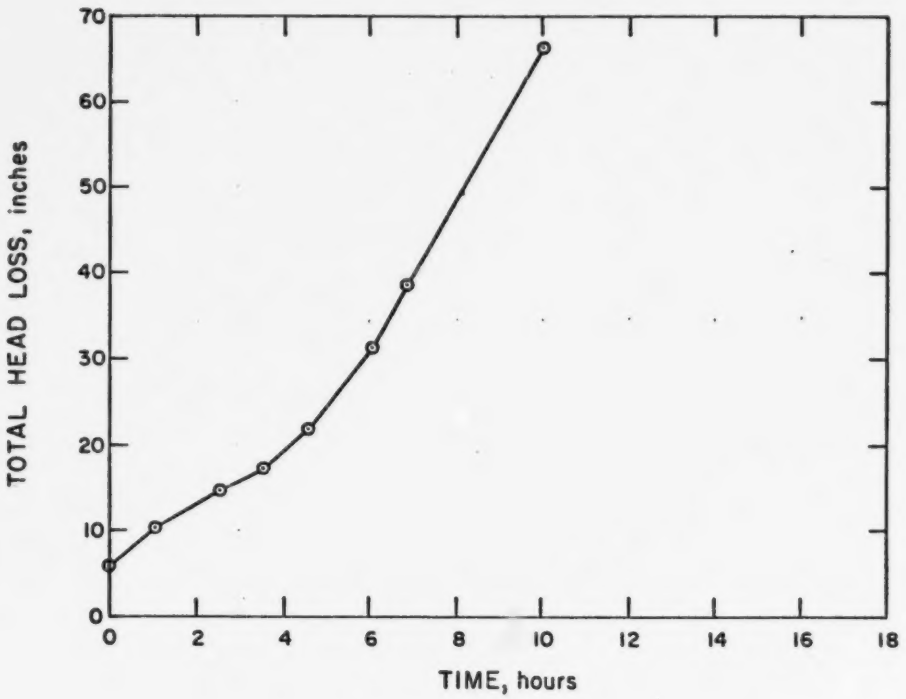
HEAD LOSS PROFILES AT TERMINAL CONDITIONS. 24.0 mg/l CLAY, 3.0 gpm/sq.ft.

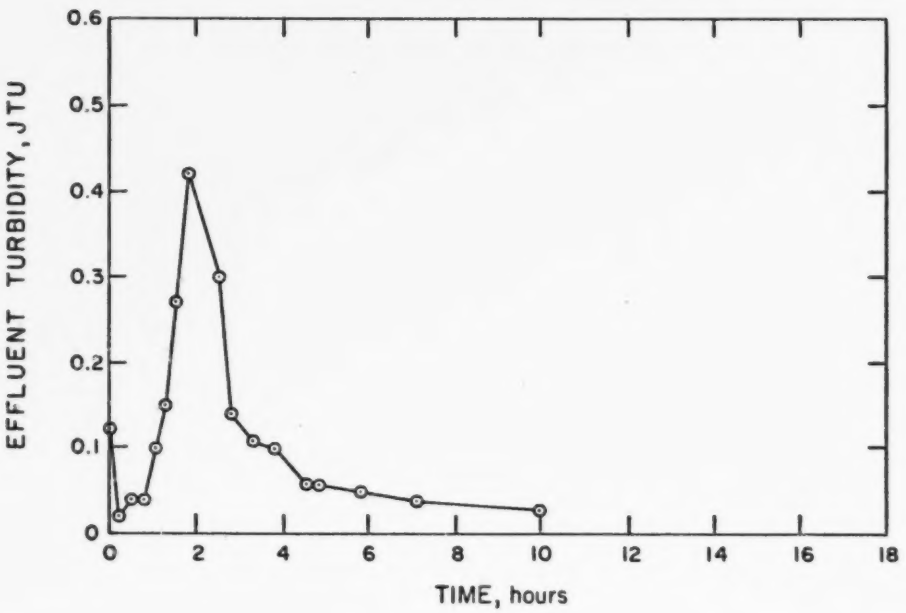




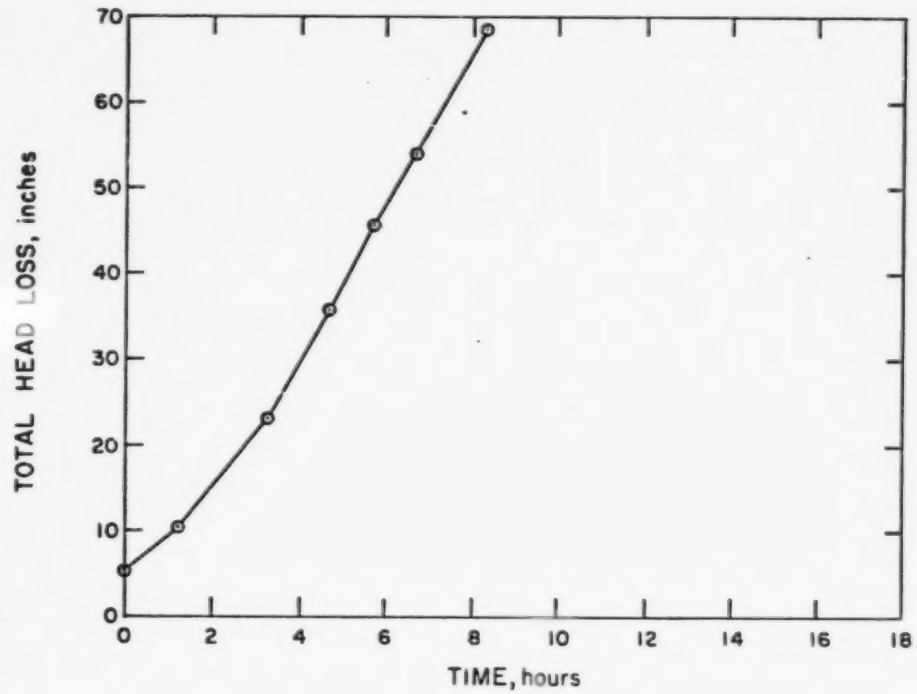


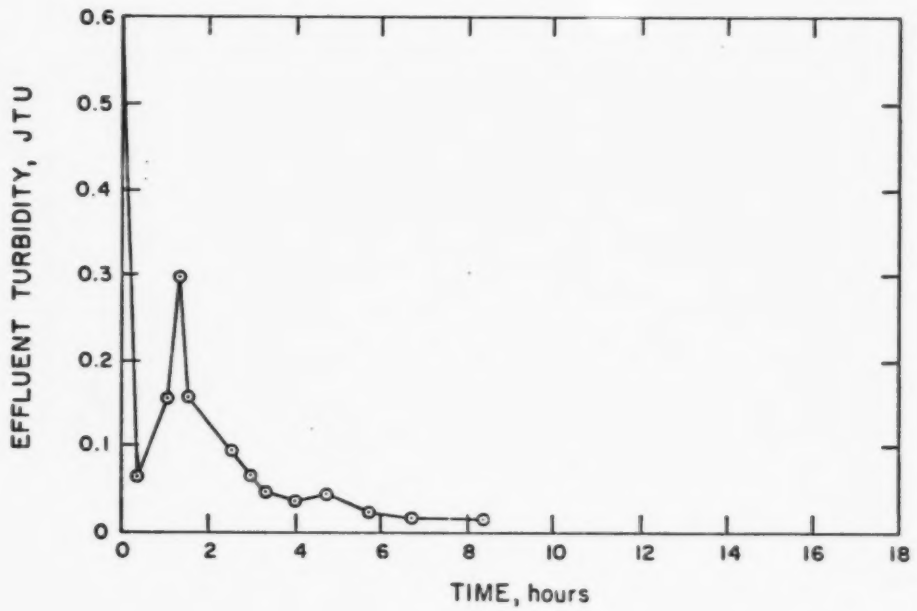
CONTACT FLOCCULATION AT 9.0 gpm/sq.ft. WITH 1.0 mg/l CAT FLOC, 6.0 mg/l CLAY, 20 mg/l ALUM FIRST TWO HOURS



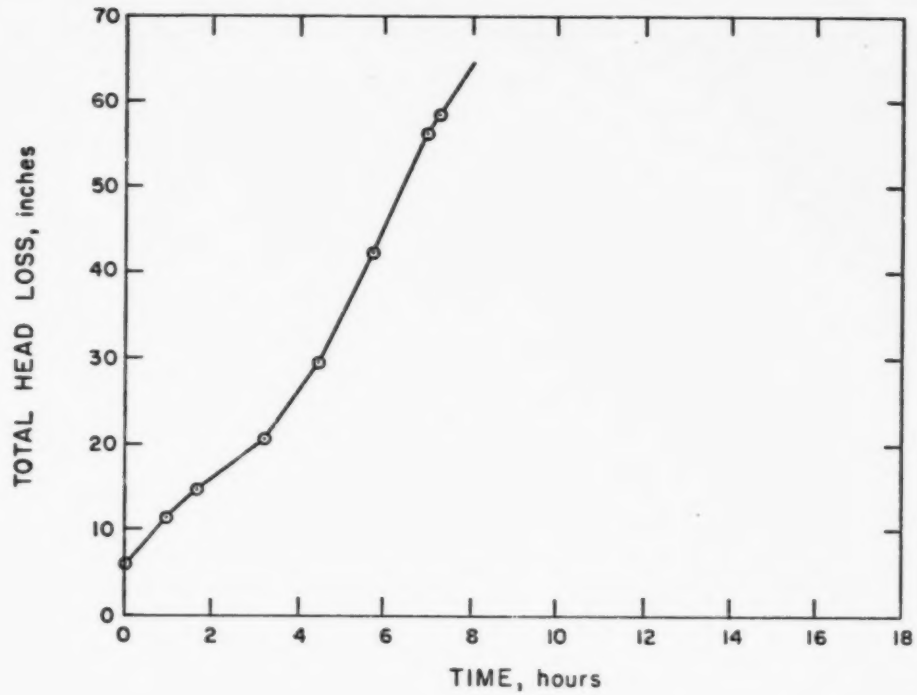


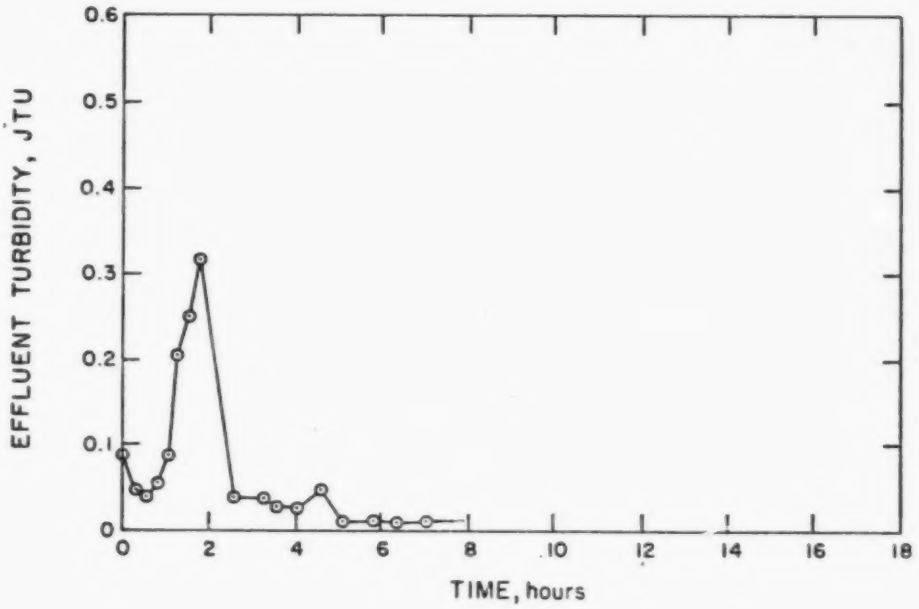
CONTACT FLOCCULATION AT 9.0 gpm/sq.ft. WITH 1.0 mg/l CAT FLOC, 6.0 mg/l CLAY, NO ALUM ADDED



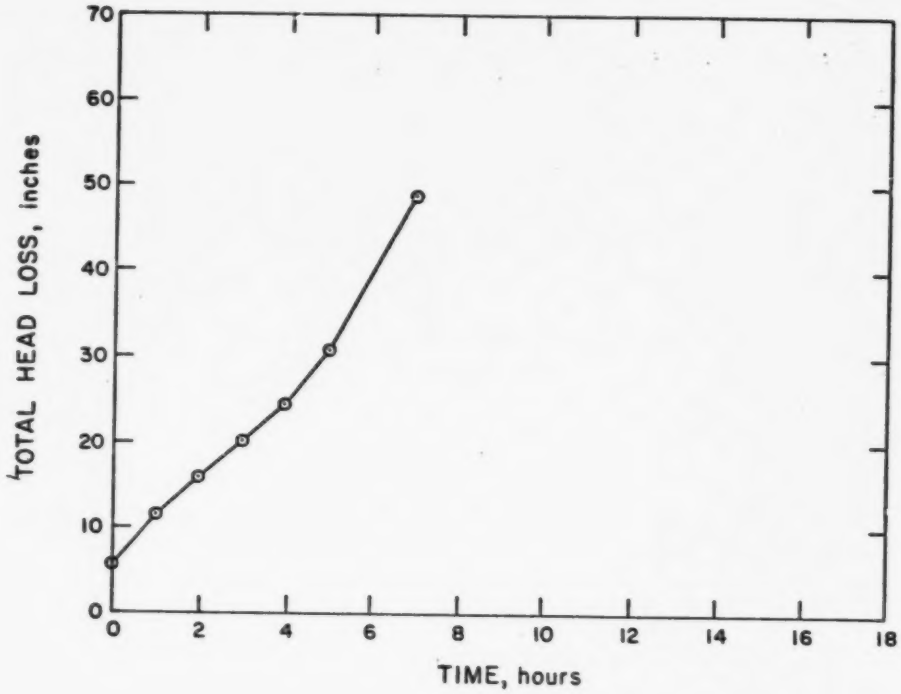


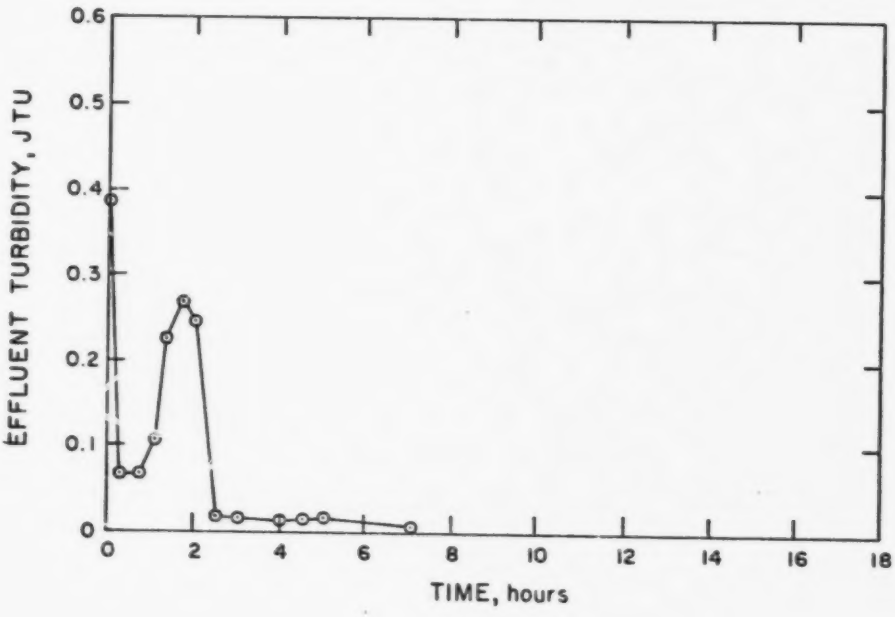
CONTACT FLOCCULATION AT 9.0 gpm/sq. ft. WITH 2.0 mg/l CAT FLOC, 6.0 mg/l CLAY, 20 mg/l ALUM FIRST TWO HOURS



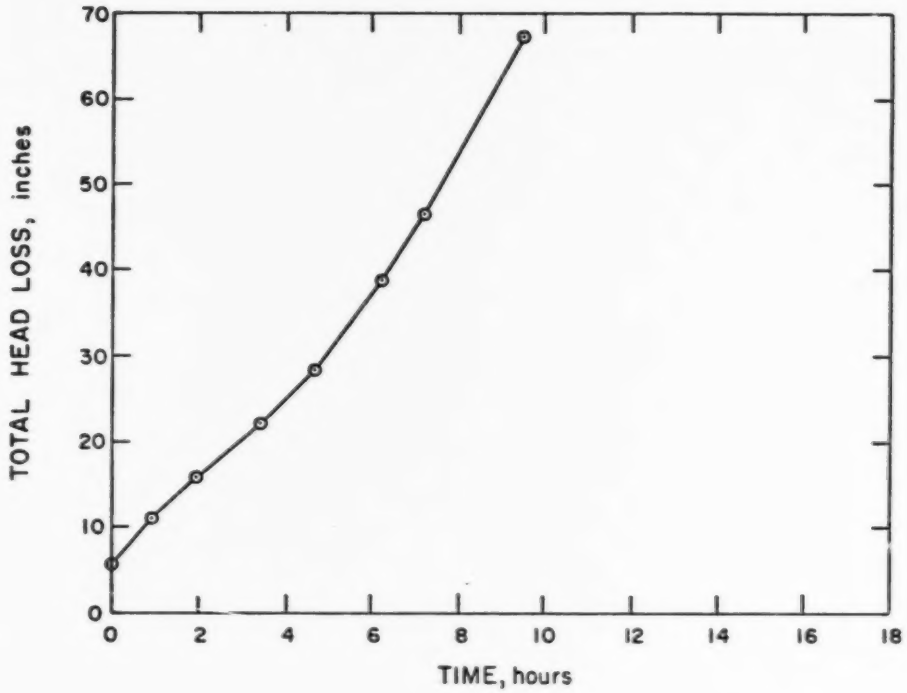


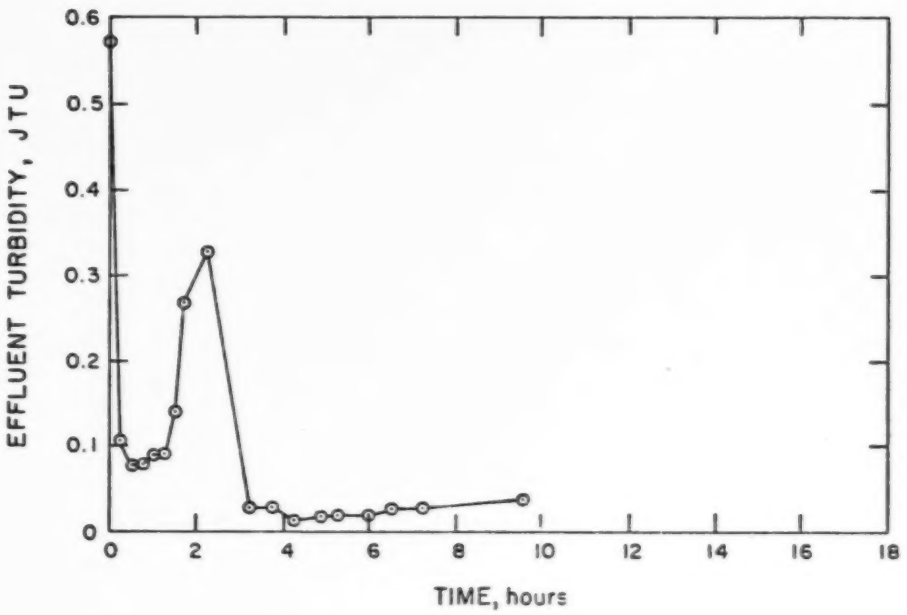
CONTACT FLOCCULATION AT 9.0 gpm/sq. ft. WITH 3.0 mg/l CAT FLOC, 6.0 mg/l CLAY, 20 mg/l ALUM FIRST TWO HOURS



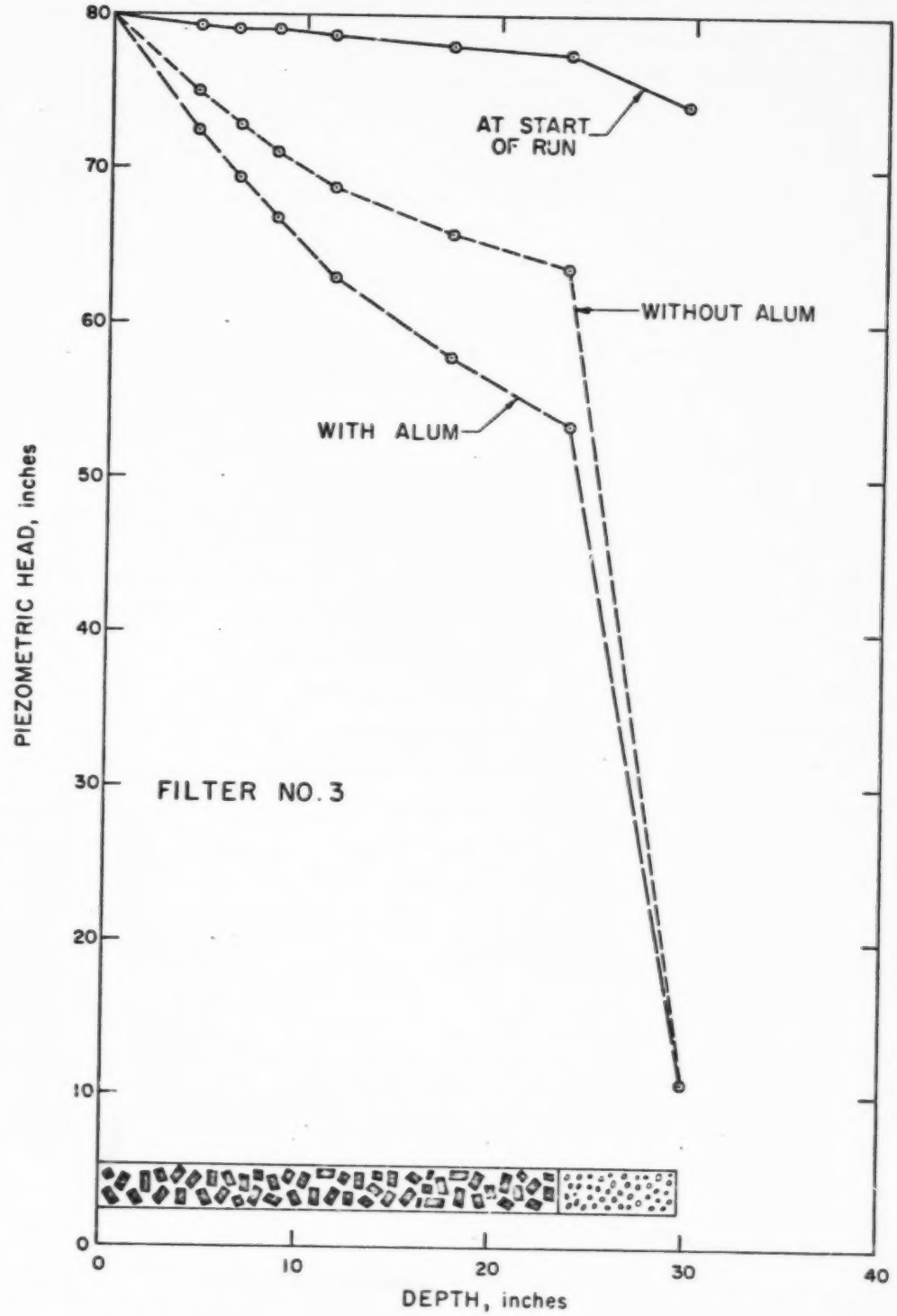


CONTACT FLOCCULATION AT 9.0 gpm/sq.ft. WITH 4.0 mg/l CAT FLOC, 6.0 mg/l CLAY, 20 mg/l ALUM FIRST TWO HOURS





EFFECT OF ALUM ON HEAD LOSS PROFILE AT TERMINAL CONDITIONS 1.0 mg/l CAT FLOC, 9.0 gpm/sq.ft.



CHAPTER IV

IMPLICATION OF RESULTS IN PROCESS DESIGN

CONTACT FLOCCULATION AND THE TRADITIONAL CLARIFICATIONS SYSTEM

The question of whether to use contact flocculation or the traditional clarification system is of importance from the practical viewpoint. As contact flocculation has not been practiced as extensively as the traditional flocculation-sedimentation-filtration system, technical data on operational experience is very scarce. The experimental results of this study indicate that contact flocculation is a feasible method to clarify turbid water and to produce a high quality product in an installation which is considerably simpler than the traditional three-stage water clarification system.

The selection of one system over the other is determined by a variety of factors and the relative importance of each factor varies according to the specific conditions under consideration. The single factor which can most significantly affect the decision is the concentration of suspended matter in the raw water. Obviously contact flocculation is most attractive for clarifying low turbidity water. As all suspended solids are removed by the filter, the length of filter run decreases with increasing suspended solids load. In the traditional three-stage system, the bulk of the suspended solids is removed in the settling unit and the length of filter run is practically independent of the raw water suspended solids. The ultimate criterion upon which a decision in favor of one system or the other can be made is the unit cost of the product. The length of filter run has a marked effect on the cost of treatment, particularly when filter runs are short. Although no general rules can be set that will apply indiscriminantly to all situations, some estimates can be made on the basis of the experimental results.

Assuming that backwashing is practiced at a rise rate of 75-in/min and lasts 10.0 minutes, the percentage of product water used for backwash can be expressed as;

$$P_{BW} = \frac{780}{T_H q} \qquad \text{(IV-1)}$$

Where T_H is the length of run in hours and q is the filtration rate in gpm/sq ft. Using this equation and the experimental data of Table III-1, a plot of percent backwash water against raw water suspended solids was

prepared and is shown in Figure IV-1. According to Koenig (Reference IV-1), 90 percent of the backwash cost is the cost of water used for backwashing. Based on Koenig's study, the total cost of water treatment in a 8 mgd three-stage plant operated at design capacity is 7.71¢/1,000 gal (1964 prices). The contribution of various cost components is given in Table IV-1.

TABLE IV-1

CONTRIBUTION OF COST COMPONENTS IN A TYPICAL PLANT

Component	¢/1,000 gal	% of Total Cost
Investment	3.30	42.8
Manpower	1.60	20.8
Energy	1.70	22.1
Chemicals	0.72	9.3
Miscellaneous	0.39	5.0
Total	7.71	100.0

The major cost difference between the traditional three-stage plant and contact flocculation will be in the investment component. With the elimination of sedimentation and flocculation units, a conservatively estimated 25 percent of the investment can be saved. Conversely, the cost of backwash will increase from the current 0.19¢/1,000 gal estimated by Koenig to a higher value which depends on the backwash percentage. Assuming that all other changes in cost are negligible, an estimate of the limiting suspended solid contents above which contact flocculation may not be economical, can be obtained. The cost of conventional treatment without flocculation, sedimentation, and filter backwashing is 6.70¢/1,000 gal (6.70 = 7.71 - 3.30 +0.75 x 3.30-0.19). Hence, the cost of contact flocculation is given by

$$C_{cf} = 6.70 + C_{BW} \qquad \text{(IV-2)}$$

where C_{cf} and C_{BW} are the total cost of treatment by contact flocculation and the cost of backwashing respectively, both in ¢/1,000 gal. The cost of backwashing alone is given by

$$C_{BW} = \frac{1.1\ C_{cf} P_{BW}}{100} . \qquad \text{(IV-3)}$$

where P_{BW} is the backwash percentage. Substituting Equation (IV-3) into Equation (IV-2) and solving for C_{cf} yields

$$C_{cf} = \frac{670}{100 - 1.1\ P_{BW}} \qquad \text{(IV-4)}$$

Solving Equation (IV-4) for P_{BW} with C_{cf} equals 7.71¢/1,000 gal gives the upper limit of the backwash percentage as 11.9 percent. Thus, under the above conservative assumptions contact flocculation will compete favorably with traditional three-stage clarification if the required backwash percentage is less than 11.9 percent.

This very approximate analysis serves to illustrate some of the arguments involved in selecting between contact flocculation and traditional clarification systems. The numerical values given or calculated here can be regarded as rough estimates or first approximation only and should be interpreted with great care. For any specific set of conditions, a complete cost study should be carried out before a decision is made.

The effect of increasing flow rate on the cost of treatment requires consideration of two opposing factors. The possible increase in cost due to greater backwash requirements (see Figure IV-1) needs to be balanced against the savings resulting from the reduction in investment. For a relatively clear raw water, the increase in flow rate from 3.0 gpm/sq ft to 9.0 gpm/sq ft involves a 33 percent increase of backwash requirements, and a 300 percent reduction in plant size. In this case, high flow rates are advantageous. For more turbid raw water, optimum flow rates may shift towards lower values. Among the few plants where contact flocculation is practiced, one reports (Reference IV-2) considerable savings by using mixed media filters at a flow rate of 4.7 gpm/sq ft with backwash requirement of about five percent.

The chemical requirements for contact flocculation are also expected to be less than in a traditional clarification system. The reason for this is that in the latter case a rather large and heavy floc must be produced in the flocculator in order for the floc to be effectively removed in the sedimentation tank. In contact flocculation large flocs are often undesirable and the chemical dose is less than the optimum for flocculation. Also, there are indications that due to compaction, the sludge carried with the backwash water is more dense than that coming from a sedimentation tank, thus making contact flocculation sludge disposal less expensive.

The operation of contact flocculators is more complicated than that of conventional rapid sand filters for several reasons. The lack of flocculation

and sedimentation facilities exposes the filters directly to the raw water, and any fluctuation in water quality that can affect filter operation and performance has to be detected and matched by an appropriate remedial action. Because the entire clarification process is carried out in a single unit, the filter, accidental failure of that unit cannot be permitted. This means that the unit has to be under constant and strict control.

CONTROL SYSTEMS FOR CONTACT FLOCCULATION

The functioning of a contact flocculator depends to a great extent on the type and dose of chemicals added to the raw water before it is applied to the filter. These, in turn, depend on the quality of the raw water, mainly the suspended solid concentrations, and also other chemical and physical parameters. If the raw water source is of fixed properties, the problem is relatively simple. Changing raw water quality may cause some difficulties in operation. The problem stems from the need to detect the change in quality and determine the remedial step before results of the change are manifested in the quality of the treated effluent. Although the early detection of a quality change may be accomplished with relative ease, the determination of an appropriate corrective measure requires some sophistication. If every individual water quality parameter had a known effect on filter performance and the effect was independent of all other quality parameters, a control system could be devised in which all parameters were monitored and corrective measures taken automatically. Because this is not the case, the only rational control system is one that analyzes continuously the overall effect of the conditioned water on the performance of the contact flocculators.

Other investigators have reported on the development of control systems for filtration with the conventional three-stage clarification system. Conley and Evers (Reference IV-3) describe a system composed of two pilot filters, operating alternately, onto which the flocculated waters are applied and the effluent turbidity monitored. Any change in floc properties can be detected within 10 minutes according to the authors. The pilot filters were not geometrically similar to the prototype, nor was the chemical conditioning of the filter influent identical to that of the prototype. Hence, the performance of the pilot filters gave some indication on the expected performance of the prototype, but no direct correlation appeared to exist between pilot and prototype filter operation. Kreissl, Robeck, and Sommerville (Reference IV-4) suggested the use of a pilot filter which was

geometrically similar to the prototype. In order to determine the adequacy of floc strength in the prototype filter without waiting for excessive clogging to develop, they proposed that the pilot filters be operated at an "equivalent flow rate" for which the pore velocity was equal to that in a clogged filter. This test determined in a relatively short time whether the terminal headloss could be attained without floc breakthrough, a condition which is necessary for optimum filter operation.

In order to develop a practical and reliable control system for contact flocculation, additional studies are required. It is thought that such a system would require a pilot filter as the primary sensing device. A careful study of the correlation between pilot and prototype performance will be necessary before the successful operation of a control system can be expected. Development of headloss profiles and floc penetration in the initial phase of a run of the pilot filter and their relation to subsequent functioning of the prototype are subjects of great concern in connection with control systems.

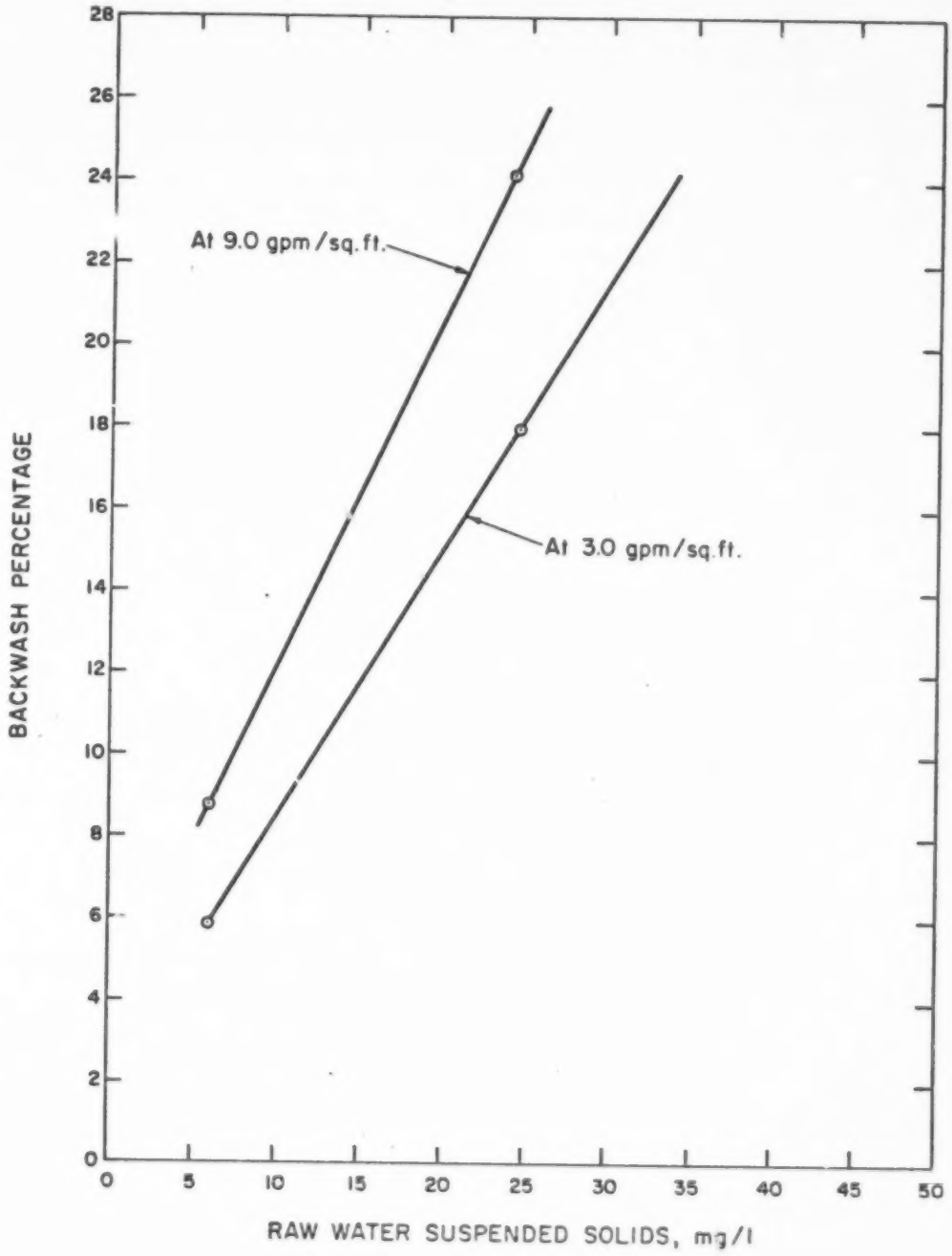
As part of the experimental work of this study a small pilot filter was operated in parallel with the pilot plant filters employed in the investigation in an attempt to establish a correlation in performance between the two. The pilot filter was a 1.0-in diameter, 6.0-in length column, which contained a very fine anthracite media. It was operated at 3.0 gpm/sq ft using the same suspension that entered the pilot plant filters. The turbidity of the effluent from the pilot filter was measured periodically. The results of two of such tests are shown in Figure IV-2. It is apparent that a pilot filter could provide, in a very short time, an indication of the performance to be expected with a large-scale filter. However, it is also clear that further work will be necessary to establish criteria for the design and operation of pilot filters.

FILTRATION AND CONTACT FLOCCULATION IN WASTE TREATMENT

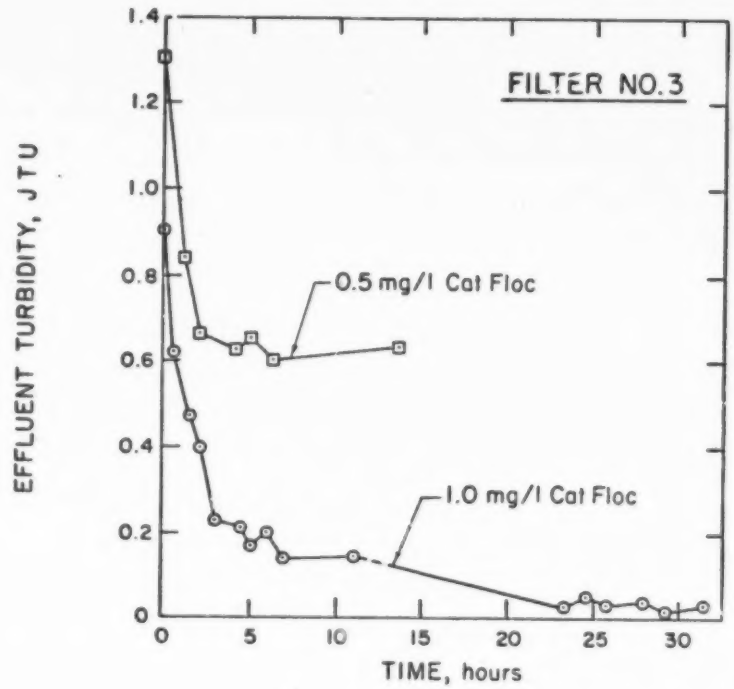
The experimental work and the discussions presented in the previous chapters are concerned with clarification processes in water treatment. The application of these processes to wastewater treatment is a subject of increasing interest. Recent trends in water pollution control have introduced a new approach to the concept of wastewater treatment. To the traditional concern for removal of suspended solids and biodegradable carbon, there has been the added problems of refractory substances, toxicants, and biostimulants and the desirability of improved clarity of effluents. The subject of

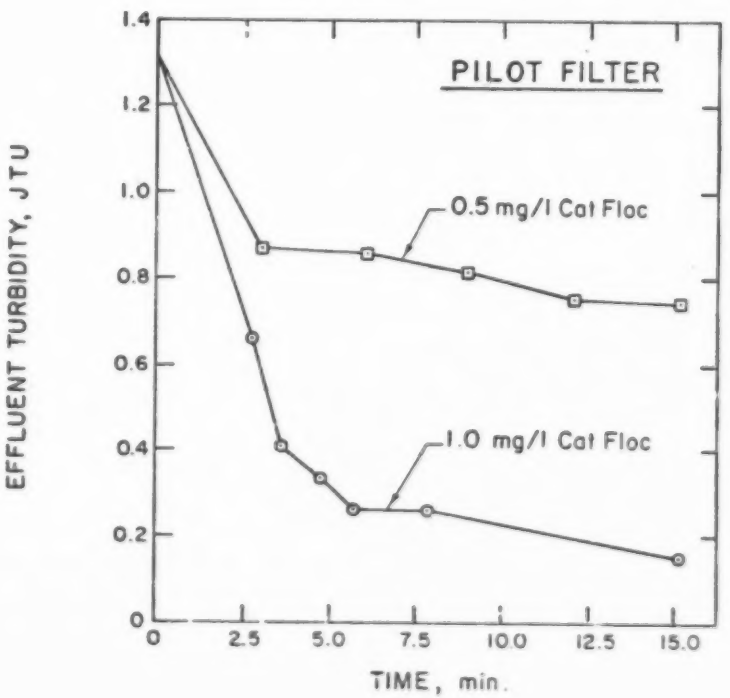
advanced waste treatment is concerned, according to Weinberger (Reference IV-5), with the development of techniques and systems for treating any type of waste to any desired level of quality. In order to achieve wastewater clarification for direct reuse or other applications requiring a very clear water, the techniques used for domestic water clarification will have to be transferred to the wastewater treatment industry. Filtration or contact flocculation are already in use for polishing secondary plant effluents. Culp and Hansen (Reference IV-6) reported on the use of mixed media filters for polishing the effluent from an extended aeration plant. With a suspended solids load of up to 120 mg/1, they were able to maintain filter runs of 15 to 24 hrs at 5.0 gpm/sq ft. In addition to the key role which contact floccuation may have in advanced wastewater treatment, this process has a great potential in many chemical wastewater treatment plants.

EFFECT OF RAW WATER SUSPENDED SOLIDS ON BACKWASH REQUIREMENTS



COMPARISON OF PILOT FILTER AND PILOT PLANT FILTER PERFORMANCE





CHAPTER V

SUMMARY AND CONCLUSIONS

Experiments were carried out on four pilot filters to study the interrelation between floc properties and filter performance. Two types of filter systems were examined. In the first, the filtration process was preceded by flocculation and sedimentation similar to the traditional water clarification system traditionally employed by the water treatment industry. In the second, the flocculation and sedimentation units were eliminated and raw water was applied to the filter after coagulant additions and a short mixing period. When operated in this latter system, the filter becomes a contact flocculator. The raw water for both phases of the study were of fixed quality adjusted to simulate a local surface source. Filter operation was characterized by measuring the headloss profiles and the effluent turbidities. The duration of filter runs was used as a measure of filter performance.

In the first phase of the study (the three-stage clarification system), the suspension was flocculated with 20 mg/l of alum and passed through the filters after settling. Floc samples were taken from the filter influent and analyzed for floc strength and mobility. A measure of the floc strength was obtained by comparing the median floc size before and after being exposed to a fixed shearing stress exerted on the floc as it passed through a capillary tube. Floc mobility was measured in a Brigg Cell. Four different filter media were tested; two were single media sand filters and two had coal-sand dual media beds. Variation in floc properties was obtained by changing the pH of the flocculated water. The following conclusions are based on the experimental results:

(1) Floc properties, filter media, and filter performance are interrelated parameters and must be considered as such in any process design.

(2) Fine media filters operate better with weak alum flocs as these flocs break and penetrate to deeper layers of the bed resulting in a longer filter run.

(3) Delay of floc breakthrough and rapid buildup of headloss are associated with fine media and strong adhesive floc.

(4) Dual media filters have a distinct advantage over the single media filters. Even a minor modification of a single media filter, namely the addition of a thin layer of coal, results in a significant improvement of filter operation.

In the contact flocculation studies, filter performance was tested using varying flow rates, chemicals, and chemical doses. Two of the pilot filters were modified replacing the media by a very coarse coal-sand dual media. Headloss profiles, effluent turbidities, and length of filter runs were recorded. The results show that contact flocculation is a technically feasible method for water clarification and in certain cases it can compete favorably with the traditional three-stage clarification system. The most important parameter in selecting between the two alternative methods is the suspended solids content of the raw water. Contact flocculation is a more economic system when the suspended solids load is low.

The experimental results have demonstrated the superior performance of coarse-to-fine dual media compared to the fine-to-coarse single media. Filter runs of up to three times longer were obtained with the coarse dual media filters. The specific conclusions include the following:

(1) Among the three types of media studied, the best performance was obtained with the coarse and uniform dual media. Lower performance was obtained with the fine well graded dual media bed. The lease desirable performance was provided by the single media sand filter.

(2) The polyelectrolyte Cat-Floc was found to be superior to alum in both technical and economic respects. The optimum dose of this chemical was proportional to the concentration of suspended solids in the raw water.

(3) Early floc breakthrough occurred in the coarse dual media filters when a low polyelectrolyte dose was applied. Increasing the dose resulted in a delay of floc breakthrough and an increase in headloss. An optimum dose was reached when headloss and breakthrough occurred at almost the same time. Headloss was always the limiting factor in the fine media filters.

(4) An increased filtration rate resulted in shorter filter runs. The total volume of filtered water in a single operation cycle appeared to decrease with flow rate.

(5) A simple conservative cost estimate showed that the raw water suspended solids concentration below which contact flocculation is more economic than the three-stage system is about 20 mg/l. This value may vary considerably if backwash techniques are modified.

(6) The development of a reliable control system for contact flocculation is the most important step in the future progress of this clarification method.

APPENDIX I-1

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APPENDIX I-2

LIST OF ABBREVIATIONS

rpm	revolutions per minute
gpm	gallons per minute
gpd/sq ft	gallons per day per square foot
mg/l	milligrams per liter
meq/l	milliequivalent per liter
gpm/sq ft	gallons per minute per square foot
in/min	inches per minute
mm	millimeters
JTU	Jackson Turbidity Units
u'sec/volt'cm	microns per second per volt per centimeter